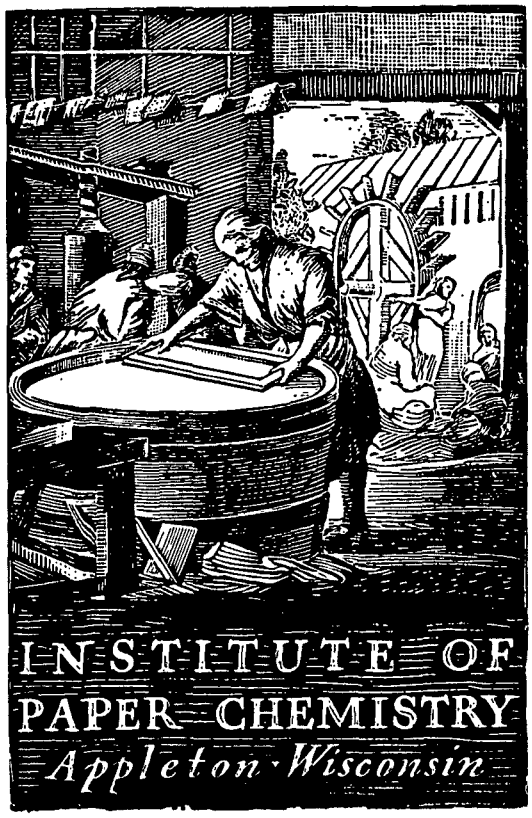

SYSTEMS ANALYSIS OF CHEMICAL AND ENERGY RECOVERY IN SULFATE PULPING

Project 2893

Report Two

A Progress Report

to

MEMBERS OF GROUP PROJECT 2893

February 12, 1971

THE INSTITUTE OF PAPER CHEMISTRY

Appleton, Wisconsin

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THE INSTITUTE OF PAPER CHEMISTRY

Appleton, Wisconsin

SYSTEMS ANALYSIS OF CHEMICAL AND ENERGY RECOVERY IN SULFATE PULPING

SUMMARY

The interim progress reports for the joint research project meeting on December 15-16, 1970 are cumulated here. In addition, the flow sheets which were supplied by the cooperators for twelve typical kraft pulp mills were reviewed to determine whether the particular equipment types which were being studied under this initial project were sufficiently common to be useful to each cooperator. In most cases, the equipment which is being studied will be directly applicable to mills which are now in operation by each of the supporting companies.

The accuracy of response of the Kamyr digester model has been significantly improved by the use of a semiempirical chemical mechanism involving the definition of two different forms of lignin with different reactivities to hydroxide and thiol groups. The resulting parallel reactions give a much better fit to the experimental pulping data than do mechanisms of equal complexity assuming series reactions of lignin moieties.

The mathematical model proposed earlier for long-tube vertical evaporators was elaborated and applied to the description of the seven-effect evaporator system at Crown Zellerbach's Wauna, Oregon mill. The steady-state heat and material balances are presented, as well as the recognized needs for proper parameter estimation to develop the dynamic model.

A detailed analysis of the control strategies and present operating conditions on the Combustion Engineering recovery furnace of Owens-Illinois, Valdosta, Georgia is given as a basis for a proposed computer control scheme, and the necessary

block diagram and functional components for the control scheme are given. A critical review of the theoretical and practical factors in the mechanisms of spray evaporation, burning, and char-bed formation are presented as a basis for the dynamic model of the active zones of the furnace. This information will be used as a basis for the development of the dynamic model of the furnace which will in turn be used to simulate the effectiveness of the control scheme which has been proposed.

The general form of the kiln equations is presented, together with data taken this summer on the temperatures and moisture contents for the rotary lime kiln of American Can Co., Halsey, Oregon. A possible mathematical method for estimating the best possible values of the mass and energy transfer coefficients for this model of the kiln is also presented. The definition of the proper variables for the evaporator and recovery furnace has been completed. In addition, the relationships for a single evaporator body have been listed. Work is proceeding on similar 'lumped' presentations of the behavior of the other components of the system for the overall, integrated, steady-state model.

INTRODUCTION

This work on a systems analysis of the chemical and energy recovery in sulfate pulping is directed toward a critical evaluation of how to get the most profit from the application of process control to the kraft recovery system. This goal is to be attained through a combination of individual model and control studies, as well as an integrated study of the steady state and dynamic behavior of the entire system. The primary progress thus far has been in developing individual dynamic descriptions of the digester, washer, evaporators, recovery furnace, and kiln.

The portion of the work being done at Purdue University is being administered through the Purdue Laboratory for Applied Industrial Control and the Purdue graduate students are working on these specific topics under Purdue Laboratory fellowships. Additional work by the Institute has included consultation on the development of accurate models in each area, and the covering in detail of one area (vacuum washing) which was not treated by the Purdue students. In addition to the individual dynamic models developed for control studies, a steady-state model of the entire system for one particular mill has also been developed.

The final study of the integrated dynamic control of the entire recovery system has been somewhat delayed but is now beginning. In the final analysis this may prove fortunate because the individual component analyses will be in much better defined form when this integrated dynamic analysis begins.

Each of these students spent the past summer actually working in a pulp and paper mill, and each of them had an opportunity to gain first-hand experience on the physical and chemical, as well as practical, factors which affect the process units which each of them is studying in this research program. The experience should be

very useful to each of the students in foreseeing the difficulties of practical installation and operation of the control systems and strategies which they might propose.

The interim reports given here plus Progress Report Number One from the Institute concerning "Preliminary Mathematical Model of a Vacuum Drum Washer" were presented and discussed at The Institute of Paper Chemistry on December 15 and 16, 1970.

REVIEW OF MILL FLOWSHEETS

In studying the dynamic control of each of the units in the recovery system it has been necessary to choose one particular type of equipment in order to get a concrete result. In consultation with the advisory committee on this project, the specific example of a so-called typical kraft mill as an initial starting point was chosen. The recovery system in this particular mill is being used to develop the steady-state model. The individual dynamic models are being developed for the same type of units which are in the selected mill. Although initially the data on the individual units will be from similar units in other mills, the basic engineering approach which is being taken should allow straightforward adjustment of the model to fit the particular units in the selected mill, or units in your own pulp mills. Although the actual values of the constants may be different for different individual units of the same type, the results of these studies should give a valid framework for convenient application to other individual units. The question does arise, however, as to those units which are not covered in the original analysis. To determine how critical a problem this might be, the individual mills supplied the Institute with general flow sheets for one or more "typical" kraft mills within their company. These flowsheets have been boiled down to a common form, the types of units used in each case have been identified, and are reviewed in the following sections.

DIGESTERS

A total of twelve flowsheets from the supporting companies were reviewed. The digester designs and arrangements which were represented are shown in Table I.

TABLE I

DIGESTERS

Mill	Type of Digester: Continuous (Kamyr) Vertical	Continuous (Kamyr) Inclined	Batch Direct	Batch Indirect
1	2	--	--	--
2	no data			
3	1	--	--	--
4	1	--	--	--
5	--	--	4	--
6	--	2	--	--
7	--	--	11	--
8	2	--	8	--
9	--	--	--	6
10	1	--	--	--
11	1	1	--	--
12	1	--	--	--

Most of the mills have at least one continuous digester, but some have batch digesters either exclusively or in addition to the continuous unit. This is perhaps the most serious gap in the entire present project.

The initial studies are all being performed on the continuous digester. It will be necessary to critically review these results when they are available to see how the chemical and physical basis of the model can be translated into the behavior of the batch digester, and what additional work will be needed in this area.

KNOTTERS, CLEANERS, REFINERS, AND SCREENS

Data are available on the specific arrangements of the knotters, pre-breakers, refiners, and screens. For the time being there is no active work on the modeling of these units. It is assumed that they can be reasonably well defined in terms of relatively simple physical separations. The importance of these components to the entire system is acknowledged; however, their relative contribution to the overall economics of the system will have to be defined by the studies in progress on the overall steady-state model.

WASHERS

The washer arrangements which are represented are shown in Table II.

TABLE II

WASHERS

Mill	Type of Washer			
	Rotary Drum	Waco	Decker	Continuous Diffusion
1	2	--	--	--
2	no data	--	--	--
3	2	1	--	--
4	--	--	1	--
5	3	--	--	--
6	3	--	--	--
7	no data	--	--	--
8	8	--	--	--
9	6	--	--	5
10	--	--	--	2
11	2	--	--	--
12	2	--	1	--

By far the most common unit is the rotary drum vacuum washer, and this is the unit which has been selected for detailed dynamic study by the Institute.

EVAPORATORS

The basic physical unit for all the mills was the long-tube vertical evaporator, but depending on the mill requirements these are put together in a wide variety of arrangements. The modular description of an individual evaporator which is being developed will allow a direct correspondence to be made between the evaporator system model and any actual arrangement in each mill. The most common general system involves five to seven effects with split feed to the first two evaporators.

RECOVERY FURNACES

The two main types of furnaces are the flat-hearth, suspended-combustion system of Combustion Engineering and the slanted-hearth, wall-combustion units of Babcock and Wilcox. In the twelve flowsheets, there were five (5) with CE units, four (4) with B & W units, and the remaining three were not specifically identified. As expected there is an equal balance between the two designs. The CE unit was chosen initially since this is the type of individual unit which is present in the one mill which was chosen for the initial, overall, steady-state modeling work.

GREEN LIQUOR CLARIFICATION, SLAKING, AND RECAUSTICIZING

Two mills did not report on the green liquor circuit or the recausticizing area. All of the remaining ten had gravity-type green liquor clarifiers and eight specifically mentioned a dregs washer. Of the ten reporting, eight had Dorr slaker-causticizers, one had an Eimco, and one was not identified. The number of agitated causticizing tanks ranged from 3 to 8 in various arrangements, and these were followed in most cases by rotary, vacuum, drum-type mud washers.

Nine of the reporting mills showed a gravity-type white liquor clarifier in addition to the usual white liquor storage, thus completing the circuit back to the digester.

LIME KILNS

All ten of the mills which showed the kiln had conventional rotary lime kilns. A summary of the available data is shown in Table III.

TABLE III

KILNS

Mill	Rotary Kilns Number	Diameter, ft.	Length, ft.
1	no data		
2	no data		
3	1	11	275
4	1	11	175
5	1	25 (?)	101
6	1	--	--
7	2	--	--
8	3	--	--
9	1	--	--
10	1	10.5	240
11	1	--	--
12	1	11	275

The rotary, gas or oil-fired kiln is the unit which was selected for detailed study of the internal dynamics and control strategy.

CONCLUSIONS

As can be seen, there is some applicability of the results of the present studies to all of the existing mills. Although there are a few obvious gaps, the general pattern is that a very large proportion of the equipment in these mills is represented in the present initial study.

The shortcomings for those types of equipment not initially represented, such as the common batch digester, may not be critical for a proper definition and use of the results of the steady-state optimization of the process, since many of the steady-state parameters may be the same, but will have different average values. The primary gap is in the dynamic control analysis of the overall system. The blow and wash cycles and dynamics of the batch process are, of course, quite different from the fluctuations expected in a continuous digester. This is a problem which must receive further critical attention.

THE INSTITUTE OF PAPER CHEMISTRY

Robert A. Holm

Robert A. Holm, Director
Division of Industrial &
Environmental Systems

KAMYR DIGESTER MODEL: PROGRESS REPORT

BY

CLIFFORD C. SMITH

December 14, 1970

1. INTRODUCTION

There are a number of good reasons for wanting to create a model of a Kamyr Digester. Of these, better understanding, better control and better operation of the process are the motivation for this work. A better understanding of the relationships can be obtained from the model and the modelling process in many ways. One way is by "playing" with the model after it is created or another is by carefully examining the mathematical and associated physical significance of the mathematical forms used in the creation of the model and their sensitivity to the physically significant parameters of the real system. This project is basically an exercise in the latter. Better control of the process can be achieved, based on an understanding of the system by either better compensation for variation or elimination, if possible, of variation in those parameters which show a high degree of sensitivity in the model. Better operation can be achieved by reducing those variations which call for operator intervention in the control scheme (i.e. control failures). This is done by giving close attention in the modelling process to those phenomena which now require more than routine operator attention in a less than total control environment.

There are a number of good reasons why modelling a Kamyr Digester is difficult. It is generally acknowledged that the pulping reaction is a complex phenomena both chemically and

physically. It is difficult to describe in a simple form due to variability of wood structure and the largely unknown chemistry. Also measurement and characterization of the real system for verification of feasibility of modelling approaches are almost unobtainable as the digester's interior is virtually inaccessible for knowledge of temperature gradients, concentration gradients, flow patterns, etc. Further, wood's physical change, as it affects flow and chemical exchange, as it is pulped is also not known. These problems are attacked by trial and error with model concepts for pieces of a total model, proving or disproving each on the basis of limited data. A single overall approach does not exist.

Based on the preceding philosophy these particular areas have been chosen for close scrutiny. First, the cooking zone with its attendant reactions. Since those reactions are temperature dependent, the temperature profile dynamics are considered. Second, the wash zone temperature, due to the dependence on it of cooking and the actual diffusion washing process is chosen. Third, the chip column dynamics are of interest due to the large operational problems associated with "hang ups". Once a reasonable and tractable model concept is found and substantially verified for each of these areas, a total model is to be put together as a final dynamic description of the Kamyr Digester.

2. CURRENT RESULTS

2.1 Cook Zone - Kinetics

Reported last May[1] were preliminary results of an attempted complex chemical description. Since then validation of such an

approach has been unsuccessful. A more simple description is more promising in its results.

Under basically the same assumptions: ie., no diffusion limitation on reaction, high enough penetration for basically homogeneous reaction, and the splitting of ether bonds in the lignin structure as the main delignification mechanism, the following model can be written.

$$L_1 + OH \xrightarrow{k_1} \text{Soluble Products}$$
$$L_2 + OH \xrightarrow{k_2} \text{Soluble Products}$$
$$L_1 + SH \xrightarrow{k_3} \text{Soluble Products}$$
$$L_2 + SH \xrightarrow{k_4} \text{Soluble Products}$$
$$C_c + OH \xrightarrow{k_5} \text{Soluble Products}$$
$$C_h + OH \xrightarrow{k_6} \text{Soluble Products}$$

where $k_i = k_{oi} e^{-E_i/RT}$

and C_c = cellulose
C_h = hemicellulose
L_1 = β aryl ether structures in lignin
L_2 = other alkyl ether structures in lignin

and it is further assumed from the structure of lignin[2] that L_1 is 70% of the lignin and L_2 is 30% of the lignin and that all Na_2S is completely hydrolyzed to NaSH and NaOH.

Using these equations a simulation was made on the PLAIC Hybrid Computer and parameter adjustment was made with fair success to describe pulping as experimentally determined by Westvaco in typical operating ranges. The results of this model along with the Westvaco data are shown in Figures 1, 2, and 3. Attempts are currently being made to verify this model from other sources of data.

2.2 Cook Zone - Temperature

Temperature profile and profile dynamics are similar to those in the wash zone and work here was postponed until wash zone experience had been obtained.

2.3 Wash Zone - Temperature

The wash zone is characterized by a countercurrent flow of fairly low temperature (relative to cooking temperature), low solids content (from first filtrate tank), and cooked chips or pulp (the exact form is not certain). By considering the flows as their water equivalents (in a heat capacity sense) with an unknown, but reasonable, heat transfer coefficient, a simple countercurrent heat exchanger evolves.

Let T_1 = the temperature of the downflow of chip and entrained liquor

T_2 = the temperature of the upflow wash liquor

x = normalized distance measured down from the inlet of the wash zone and ending at the wash flow inlet

k = normalized heat transfer coefficient

v = water equivalent flow rate

Then:
$$\frac{\partial T_1}{\partial t} = k\,(T_2-T_1) - v_1 \frac{\partial T_1}{\partial x}$$

$$\frac{\partial T_2}{\partial t} = k\,(T_1-T_2) - v_2 \frac{\partial T_2}{\partial x}$$

at steady state $\frac{\partial T_i}{\partial t} = 0$ then
$$\frac{\partial T_i}{\partial x} = \frac{k}{v_1}\,(T_2-T_1)$$

$$\frac{\partial T_2}{\partial x} = \frac{k}{v_2}\,(T_1-T_2)$$

with $T_1(0) = T_{10}$ and $T_2(1) = T_{21}$. T_{10} and T_{21} being the inlet temperature conditions of the pulped chips and wash liquor respectively.

This Two Point Boundary Value Problem can be solved in a number of ways. A simulation on an analog computer with operator adjustment of $T_2(0)$ to meet the boundary conditions gives much insight into the problem. Figure 4 shows the drastic change in temperature profile caused by doubling the wash flow with $\frac{k}{v_1} = 20$ (i.e. for Fig. 4, ΔT of 350°F or for v_1 = 100 GPM = 752 lbs/min therefore k = 15040 $^{BTU}/_{min\ °F}$ per unit length.) In evaluation of whether or not this is a reasonable coefficient it must be remembered that the specific area of contact per unit length between pulp and wash liquor is included in this constant.

2.4 Wash Zone - Washing

Since it is becoming more and more common to add cold filtrate to the quench zone, the inlet temperature to the wash zone is much reduced and possible arguments for a constant wash zone temperature can be made (enough cold filtrate added to make wash zone inlet temperature of the pulp equal to wash heater temperature). Then the effects of the large sensitivity to the wash flow will be eliminated. In this case of constant wash zone temperature profile the equations for solids diffusing from the entrained liquor to the wash liquor are identical in form to those of the wash zone temperature profile with temperature becoming solids content and k becoming the diffusion rate constant. v_1 and $\bar{v}_2$ are now based on solids carrying capacity and refer to entrained liquor and wash liquor respectively. Figures 5 and 6 show the variation in the steady state profile of solids content for $\frac{v_2}{v_1} = 1, 2$, and 4 with $\frac{k}{v_1} = 4$ and

8 respectively. The resulting wash efficiency as defined by $(1 - \frac{\%Out}{\%In}$ as a function of wash flow is shown in Figure 7.

2.5 Chip Column Dynamics

The residence time in the digester is affected by the chip column velocity which is basically moved by gravity. The other forces at work are the fluid flow through the chip column, most notably the wash flow. Also friction caused by chips caught in screens exists. This phenomenon has not yet been modelled. However, once the chip column velocity, v(t), is determined, either just measured or controlled, the residence time of a currently emerging element of pulp can be calculated.

$$L = \int_{t-t_r}^{t-} v(\tau)d\tau$$

or $$\frac{dt_r}{dt} = 1 - \frac{v(t)}{v(t-t_r)}$$

Either equation can be used to take a tabulation of v(t) kept for the maximum possible t_r and integrate the tabulation backwards from t. After the integral equals L then the time of integration equals t_r.

3. FURTHER WORK

It is anticipated that further refinement of the kinetics model by the inclusion of Na_2S hydrolysis and by using a global parameter estimation scheme (such as a hyperbolic random search) will result in a better description. Also further verification over a wider range of conditions is required.

Generation of a cocurrent heat transfer model for the upper zone of the digester to obtain the temperature profile for cooking is now in order.

Investigation of the dynamic behavior of the temperature profiles in the wash zone is to be considered. It may prove useful as a quick acting control much closer to the blow line and hence less delay in a feedback loop. Wash efficiency with a variable wash temperature is also to be included so the interaction of cooking and washing can be studied.

Finally an adaptation of fluid flow in a porous media will be applied to the problem of chip column dynamics for use in operational considerations.

REFERENCES

1. Smith, Clifford C., "Preliminary Pulping Mechanisms Study" PLAIC, May, 1970.

2. Rydholm, S. A., Pulping Processes, Interscience 1965.

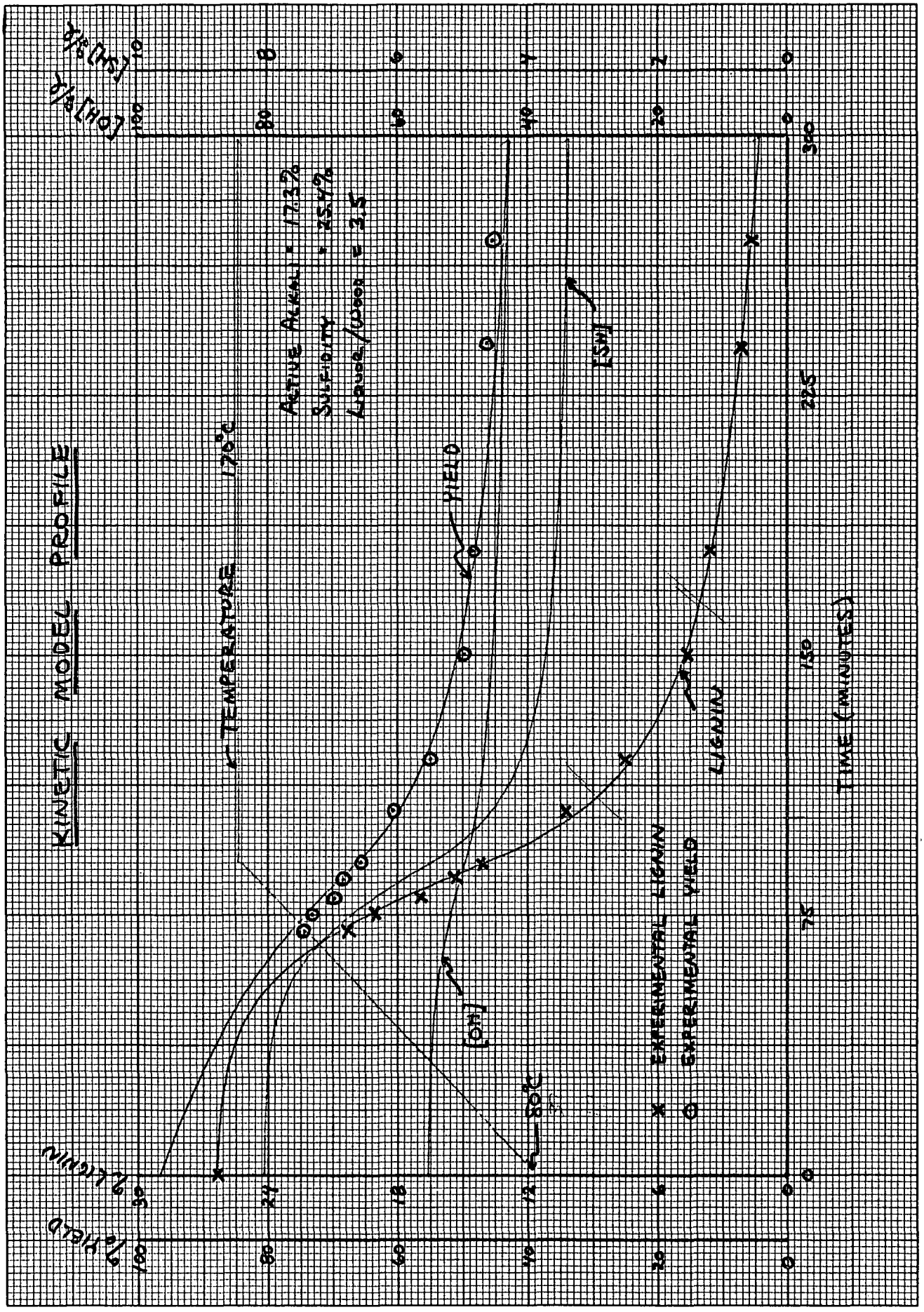
KINETIC MODEL PROFILE
TEMPERATURE
YIELD
ACTIVE ALKALI = 17.3%
SULFIDITY = 25.7%
LIQUOR/WOOD = 3.5
[OH]
EXPERIMENTAL LIGNIN
EXPERIMENTAL YIELD
LIGNIN
TIME (MINUTES)
% LIGNIN
% YIELD
[OH] g/l
176°C
80°C

FIGURE 1

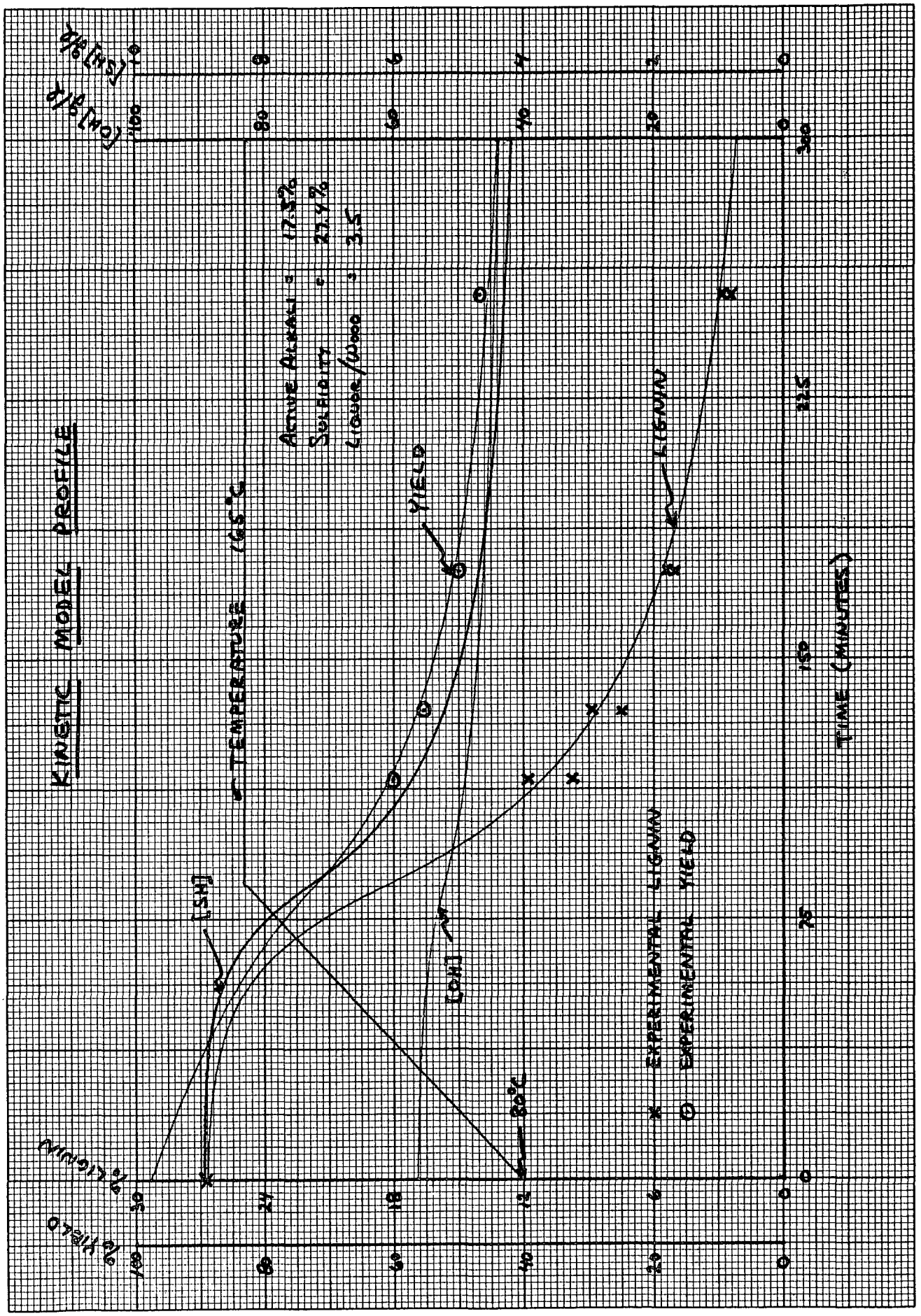
KINETIC MODEL PROFILE
Active Alkali = 17.5%
Sulfidity = 23.4%
Liquor/Wood = 3.5
YIELD
LIGNIN
[SH]
[OH]
80°C
EXPERIMENTAL LIGNIN
EXPERIMENTAL YIELD
TIME (MINUTES)
% Lignin
% Yield
% [SH]
% [OH]

FIGURE 2

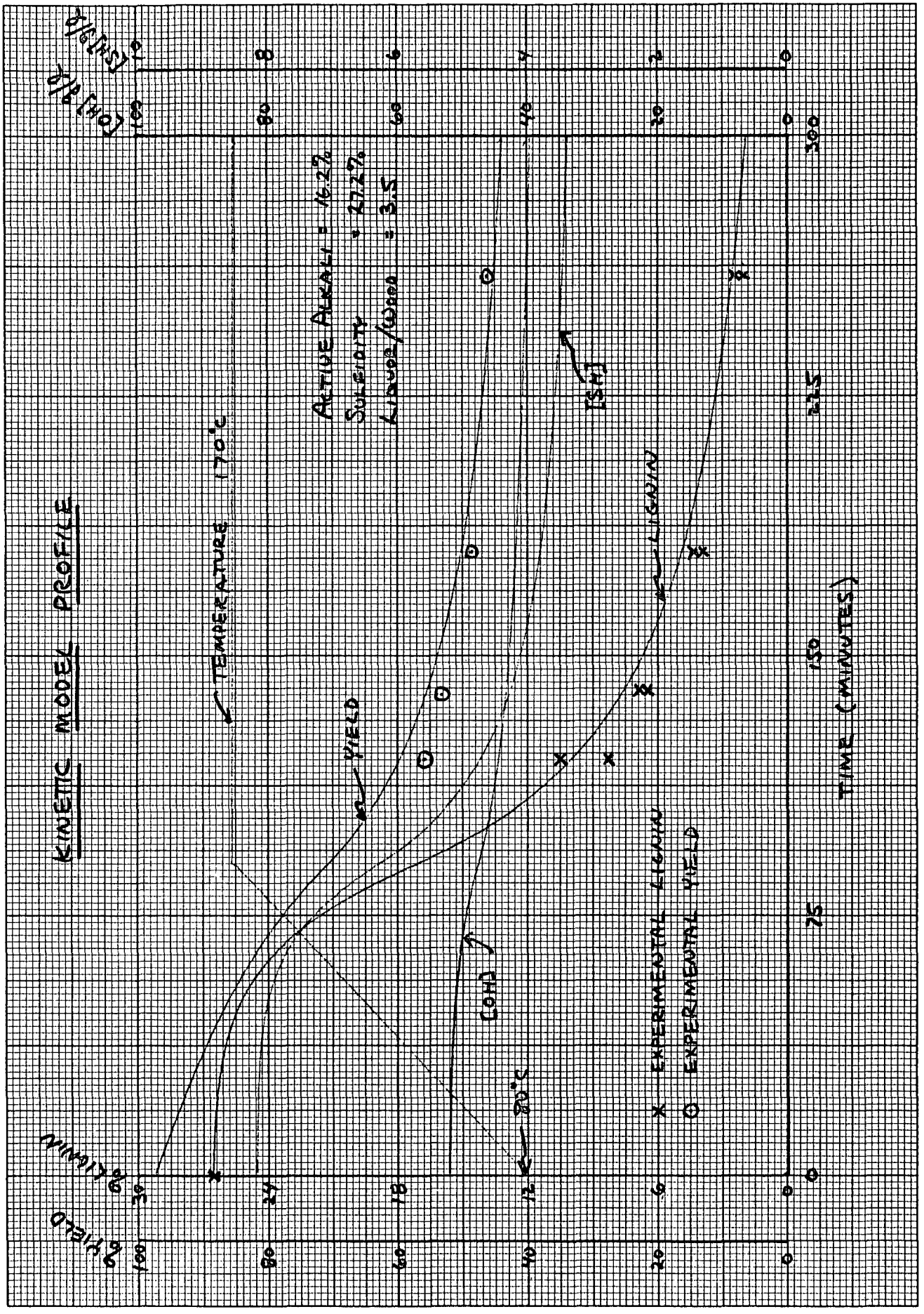
KINETIC MODEL PROFILE
ACTIVE ALKALI = 16.2%
SULFIDITY = 27.2%
LIQUOR/WOOD = 3.5
TEMPERATURE 170°C
YIELD
[SH]
[OH]
80°C
EXPERIMENTAL LIGNIN
EXPERIMENTAL YIELD
% YIELD
% LIGNIN
% (SH) & % (OH)
TIME (MINUTES)
100
80
60
40
30
24
18
12
6
0
8
4
2
0
75
150
225
300

FIGURE 3

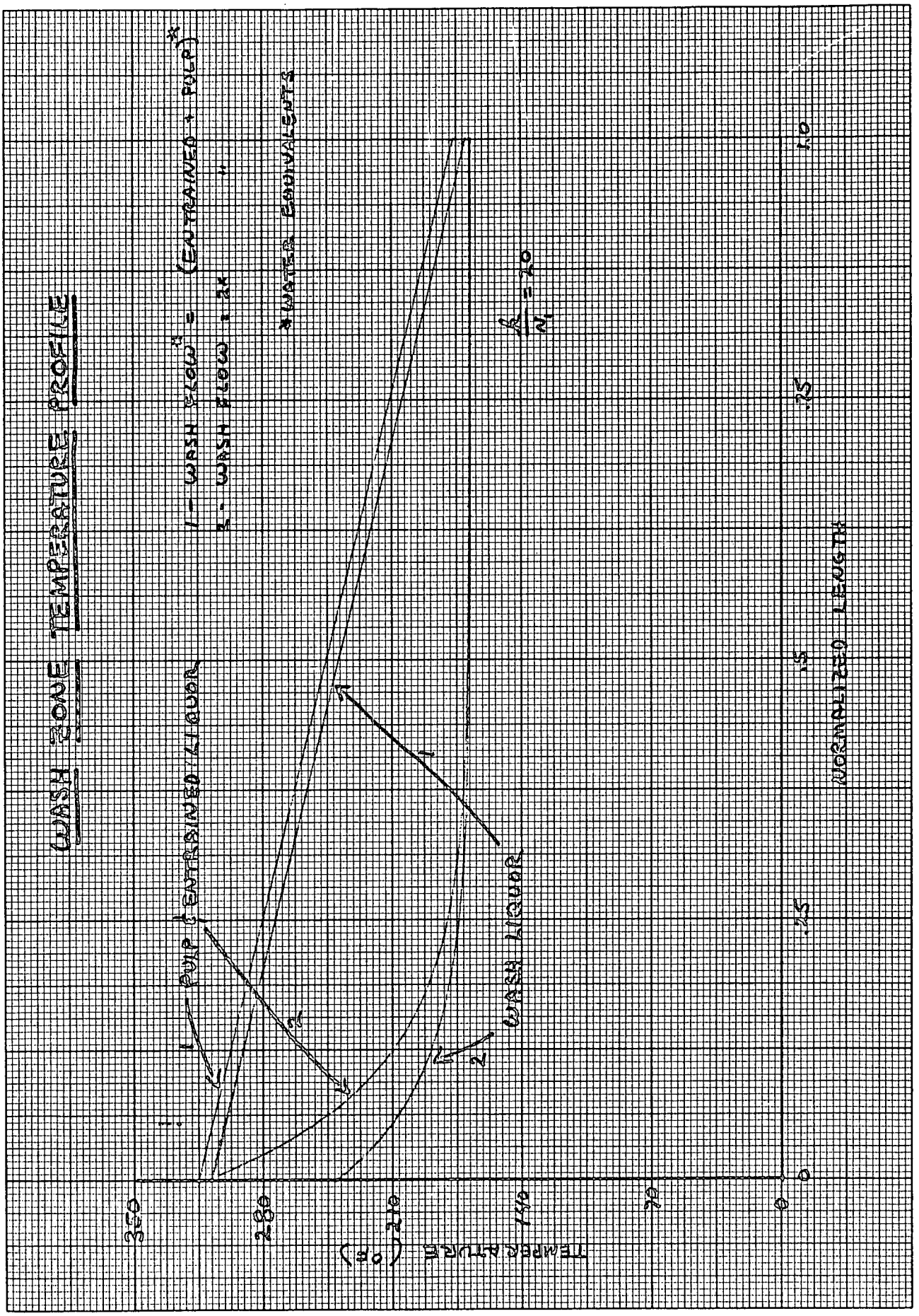
WASH ZONE TEMPERATURE PROFILE
1 — PULP ENTRAINED LIQUOR
2 — WASH LIQUOR
1 — WASH FLOW = (ENTRAINED + PULP)*
2 — WASH FLOW = 2*
* WATER EQUIVALENTS
R/N = 20
TEMPERATURE (°F)
NORMALIZED LENGTH
350
280
210
140
70
0
0
.25
.5
.75
1.0

FIGURE 4

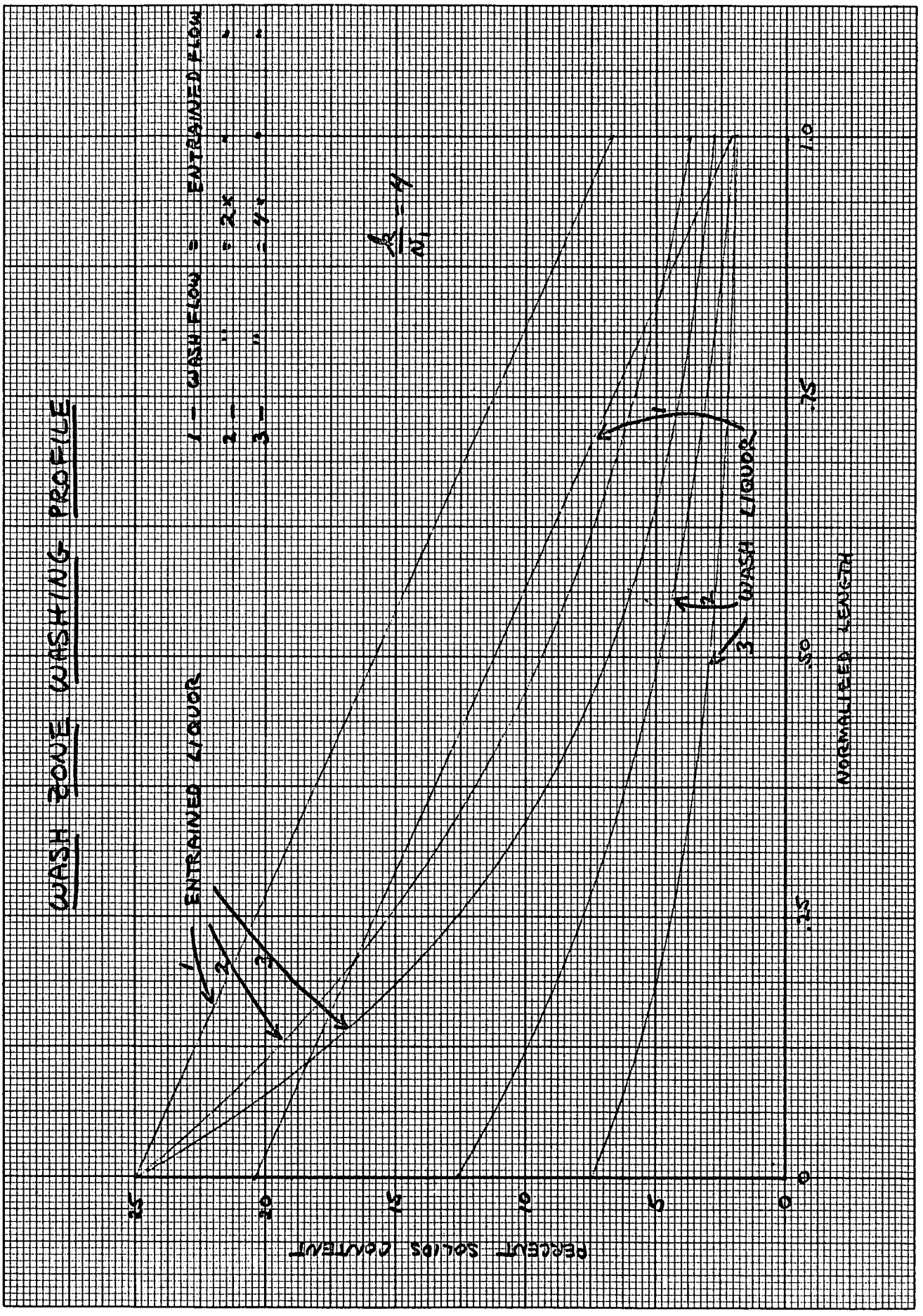
WASH ZONE WASHING PROFILE
1 - WASH FLOW = ENTRAINED FLOW
2 - " = 2x " "
3 - " = 4x " "
ENTRAINED LIQUOR
1
2
3
4
WASH LIQUOR
PERCENT SOLIDS CONTENT
35
30
25
20
15
10
5
0
.25
.50
.75
1.0
NORMALIZED LENGTH

FIGURE 5

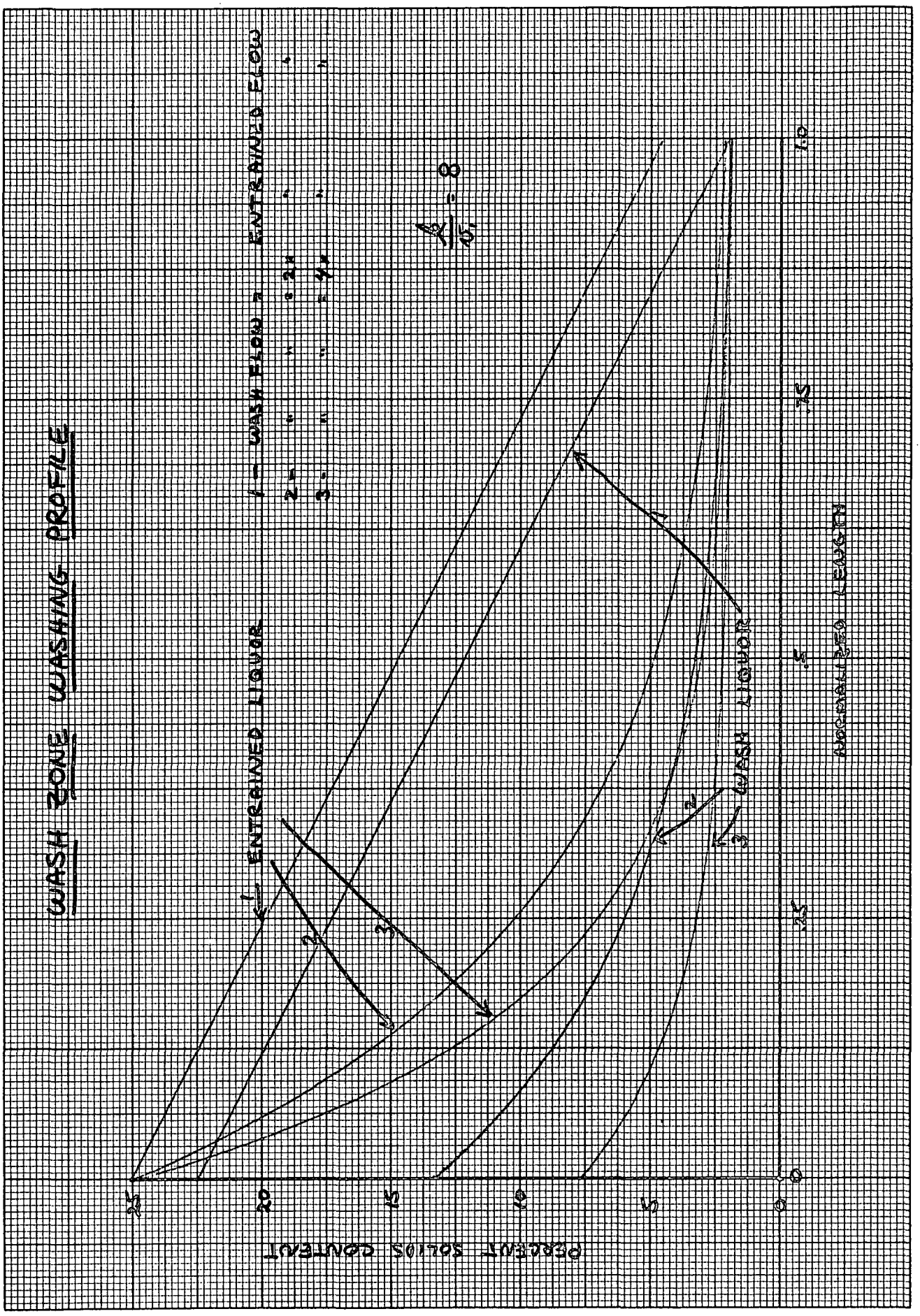
WASH ZONE
WASHING PROFILE
1 — WASH FLOW = ENTRAINED FLOW
2 " " = 2 "
3 " " = 3 "
1 ENTRAINED LIQUOR
2
3
3 WASH LIQUOR
2
1
PERCENT SOLIDS CONTENT
ACCUMULATED LENGTH
25
20
15
10
5
0
25
15
10
5
0

FIGURE 6

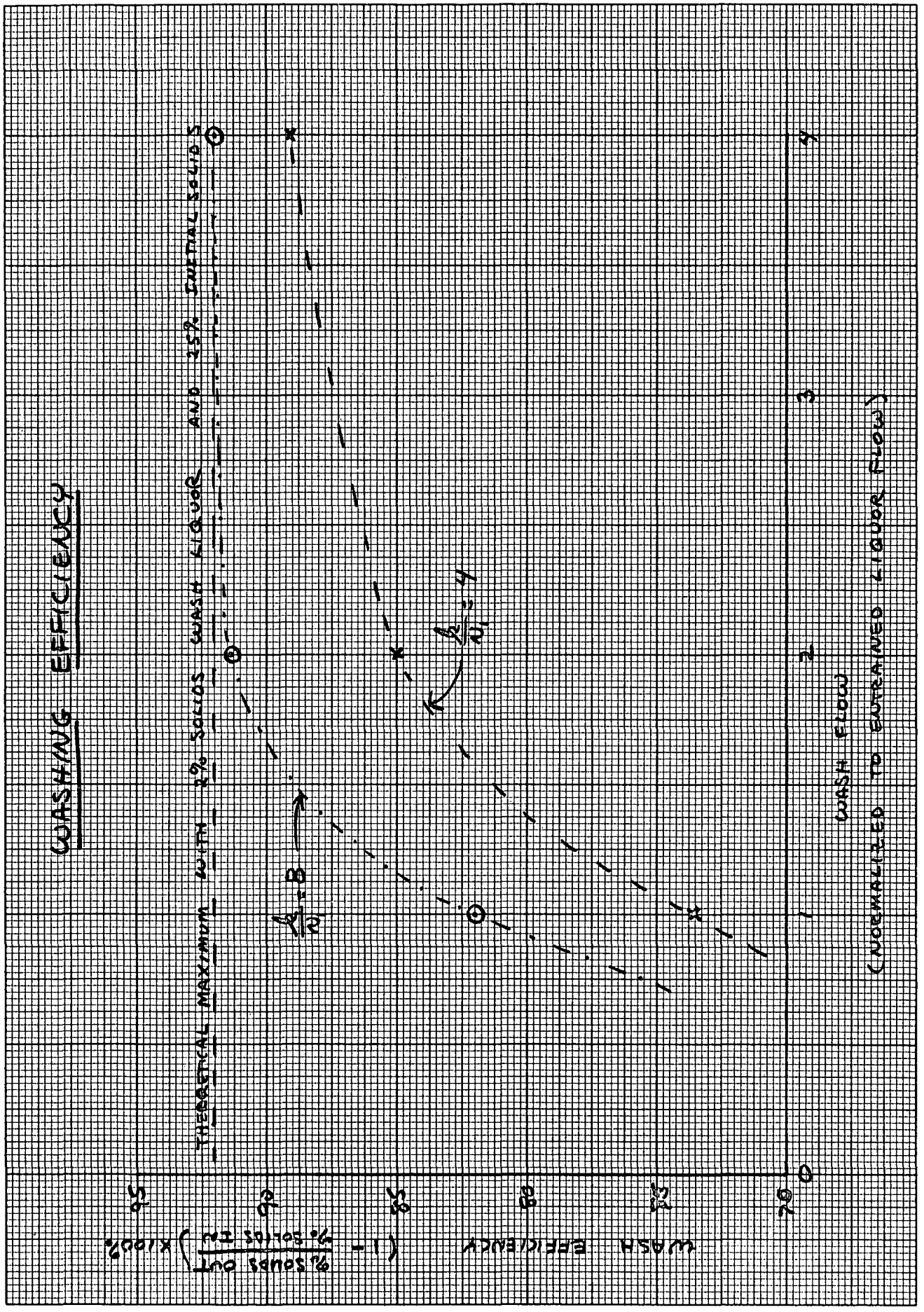
WASHING EFFICIENCY
THEORETICAL MAXIMUM WITH 2% SOLIDS WASH LIQUOR AND 25% CAUSTIC SOLIDS
WASH EFFICIENCY
(1 - % SOLIDS OUT / % SOLIDS IN) x 100%
WASH FLOW
(NORMALIZED TO ENTRAINED LIQUOR FLOW)
100
98
96
94
92
0
1
2
3
4
R2 = B
R1
R2 = A
R1

FIGURE 7

MODELLING OF MULTIEFFECT EVAPORATOR SYSTEM

BY

MICHAL V. KOTIK

December 14, 1970

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1. INTRODUCTION:

The multieffect evaporator system as a substantial part of a Kraft mill recovery area is modelled separately from the steady state over-all model. Basic data were sampled on a system at the Wauna, Oregon, Mill of the Crown Zellerbach Corporation in the summer of 1970.

Values of heat transfer coefficients and heat flows that were evaluated as a part of the Wauna evaporator system study are assumed to be valid for steady state conditions (2). The described system consists of 7 effects and no soap skim tank. The weak black liquor feed input is to the No. 4 and No. 5 effects. Enclosed is an example of a basic data set in terms of the computer program results of heat balance calculations and also in a Table on page 15.

2. THEORETICAL DESCRIPTION OF A DYNAMIC MODEL

Generally, equations describing LTV evaporator system are based on F. Galtung's outline given in Report No. 23 (1) with minor changes due to the physical parameters and the piping specifications of the Wauna septuple effect set. Hence, as was already presented (7), each effect is divided for modelling purposes into three basic sections:

A. Steam chest

B. Non-boiling section

C. Boiling section

Basic assumptions made in the Dynamic model are as follows:

1. The non-boiling section is a perfectly mixed volume.
2. The longitudinal temperature gradients are neglected in the tubes of the non-boiling section.
3. Material and energy balances for liquor in the boiling section are adequately described by algebraic equations.
4. The pressure throughout the boiling section equals the pressure in the vapor space. Therefore it is also assumed that the boiling point rise is constant for the whole section
5. The steam in the steam chest is saturated, i. e., the steam superheat is neglected. Consequently the energy of the vapor equals the energy of saturated vapor at that pressure.

6. Losses of vapor in the connecting piping are neglected.

A. STEAM CHEST

Energy balance:

$$W_{ST}H_{ST}-W_CH_C=Q_{SB}+Q_{SN}+Q_L \quad (1)$$

(for a nomenclature listing see page 28)

Steam temperature prediction:

$$T_{ST}=\frac{W_{ST}(H_{ST}+32)+U_{ST}T_{TB}(A_B+A_N)-Q_L}{W_{ST}+U_{ST}(A_B+A_N)} \quad (2)$$

Heat flows:

$$Q_{ST}=Q_{SN}+Q_{SB} \quad (3)$$

$$Q_{SB}=U_{ST}A_B(T_{ST}-T_{TB}) \quad (4)$$

$$Q_{SN}=U_{ST}A_N(T_{ST}-T_{TB}) \quad (5)$$

A typical linearization scheme example shows a derivation of a time constant τ_{ST}:

Material balance:

$$V_{ST}\frac{d\rho_{ST}}{dt}=W_{ST}-W_C \quad (6)$$

Energy balance:

$$V_{ST}\frac{d(\rho_{ST}H_{ST})}{dt}=W_{ST}H_{ST}-W_CH_C-Q_{ST}-Q_L \quad (7)$$

From equations 6 and 7:

$$\rho_{ST}\frac{dH_{ST}}{dt}+(H_{ST}-H_C)\frac{d\rho_{ST}}{dt}=W_{ST}(H_{ST}-H_C)-Q_{ST}-Q_L \quad (8)$$

Equation 4 to Equation 8:

$$\rho_{ST}=\alpha T_{ST}+\beta \quad (9)$$

Assuming that enthalpies of saturated steam and water may be approximated by

$$H_{ST}=1066+0.4T_{ST} \quad (10)$$

$$H_C=-32+T_{ST} \quad (11)$$

Equation 8 in Equation 9 gives:

$$f = \frac{dT_{ST}}{dt} = \frac{W_{ST}(H_{ST}+32-T_{ST})-U_{ST}A_B(T_{ST}-T_{TB})-U_{ST}A_N(T_{ST}-T_{TN})-Q_L}{V_{ST}(0.4\rho_{ST}+\alpha(1098-0.6T_{ST}))} \tag{12}$$

Then a linearization by Taylor series expansion about the operating point is performed, using the definition of the time constant where the time constant of a linearized equation is defined as the ratio of the coefficient of the first order term to the zero order term.

$$\tau_{ST} = -\frac{1}{\frac{\partial f}{\partial T_{ST}}}\Bigg|_{T_{STO}} \tag{13}$$

According to Andre (8) the time constant takes a final form:

$$\tau_{ST} = \frac{V_{ST}(0.4\rho_{ST}+\alpha(1098-0.6T_{ST}))}{W_{ST}+U_{ST}(A_B+A_N)}\Bigg|_{T_{STO}} = 7.8 \times 10^{-3} \text{ min} \tag{14}$$

then using the actual operating data (7) this time constant has a numerical value 7.8×10^{-3} min which is small compared to the dynamics of the rest of the evaporator system. Using a similar linearization scheme, a time constant for the tubes in the non-boiling section is found to be 4.3×10^{-2} min. and the time constant for the liquor in the non-boiling section is 0.48 min.

Consequently, Equations 6 and 7 are substituted with the corresponding algebraic equation, Equation 1 and Equation 2 can be derived.

B. Non-boiling section

1. Tubes:

$$Q_N = U_N A_{NM} (T_{ST} - T_N) \tag{15}$$

The linearization yields a time constant

$$\tau_T = \frac{m_{TN} C_{PT}}{U_{ST} A_N + U_T A_{iN}} = 4.3 \times 10^{-2} \text{ min} \tag{16}$$

2. Liquor:

Material balance:

$$\text{Solids: } V_N \frac{d\rho_S}{dt} = W_{Si} - W_{SO}, V_N = F(L_N) \tag{17}$$

Water:

$$V_N \frac{d(\rho_S \cdot X_W)}{dt} = W_{SI} X_{WI} - W_{SO} X_W \tag{18}$$

Energy balance:

$$V_N \frac{d}{dt} (\rho_\ell H_\ell) = W_{S1}(1+X_{W1}) H_{\ell 1} - W_{SO}(1+X_W) H_\ell + Q_N \tag{19}$$

Enthalpy of black liquor is approximated as:

$$H_\ell = H_{\ell W} + C_{P\ell} T_N \tag{20}$$

and

$$C_{P\ell} = 0.99 + 8 \times 10^{-5} T_N - (0.639 - 6.4 \times 10^{-4} T) X_{S\ell} \tag{21}$$

for a range T_N(100-250)°F, and

X_{SL}(5-100)% according to Harwin and Brown.

C. BOILING SECTION:

Material balance

Solids: $$W_{SI} = W_{SO} = W_S \tag{22}$$

Water: $$W_S X_{WI} = W_S X_{WO} + R_W \tag{23}$$

Generally: $$W_S = W_L X_S \tag{24}$$

and $$W_W = \frac{1-X_S}{X_S} \tag{25}$$

Energy Balance:

$$W_S(1+X_{WI})H_{\ell 1} + Q_B = W_S(1+X_{WO})H_{\ell O} + R_W H_W \tag{26}$$

The temperature of a liquor in a boiling section is:

$$T_B = BP + BPR \tag{27}$$

where BPR is determined using Dühring lines

then $$Q_B = U_B A_{BM}(T_{ST} - T_B) \tag{28}$$

The Total pressure drop:

briefly, the sum of all the pressure drops must be equal to the total pressure drop between tube inlet and exit. The computation is used for an estimation of L_N - non-boiling zone length.

$$\left(\frac{\Delta P_{FN}}{\Delta L} + \frac{\Delta P_{HN}}{\Delta L}\right) L_N + \left(\frac{\Delta P_{F_B}}{\Delta L} + \frac{\Delta P_{H_B}}{\Delta L}\right)(L - L_N) +$$

$$+ \Delta P_A = \Delta P_{TOT} \tag{29}$$

Note:

At this point it will be necessary to define a relationship between a total heat transferred and a length of non-boiling zone or a ratio of lengths of non-boiling and boiling zones. As this ratio varies in consequence of the tube scaling (fouling), also heat transfer coefficients will change. The relationship is important information for a design of the over-all steady state model. Consequently, V_N, is to be treated not as a constant but rather a function of L_N. One of the possible methods to express a change of a heat transfer rate or in other words a change of boiling zone heat transfer area due to a tube scaling is shown below:

$$Q_N = C_P(T_{Bi} - T_{Bi+1})W_{LN} = A_N U_N \Delta T \qquad (29)$$

and from (29) A_N is: $$A_N = \frac{C_P(T_{Bi} - T_{Bi+1})W_{LN}}{U_N \Delta T} \qquad (30)$$

Time derivative of (30):

$$\frac{dA_N}{dt} = \frac{d}{dt}\left[\frac{C_P(T_B - T_{Bi-1})W_{LN}}{U_N \Delta T}\right] = K\frac{d}{dt}\left[\frac{u(t).v(t)}{w(t)}\right]$$

where:

$$K = \frac{C_P}{U_N}$$

$$u(t) = (T_{Bi} - T_{Bi-1})$$

$$v(t) = W_{LN}$$

$$w(t) = \Delta T = (T_{ST} - T_B)$$

where:

$$K = \frac{C_P}{U_N}$$

$$u(t) = (T_{Bi} - T_{Bi+1})$$

$$v(t) = W_{LN}$$

$$w(t) = \Delta T = (T_{ST} - T_B)$$

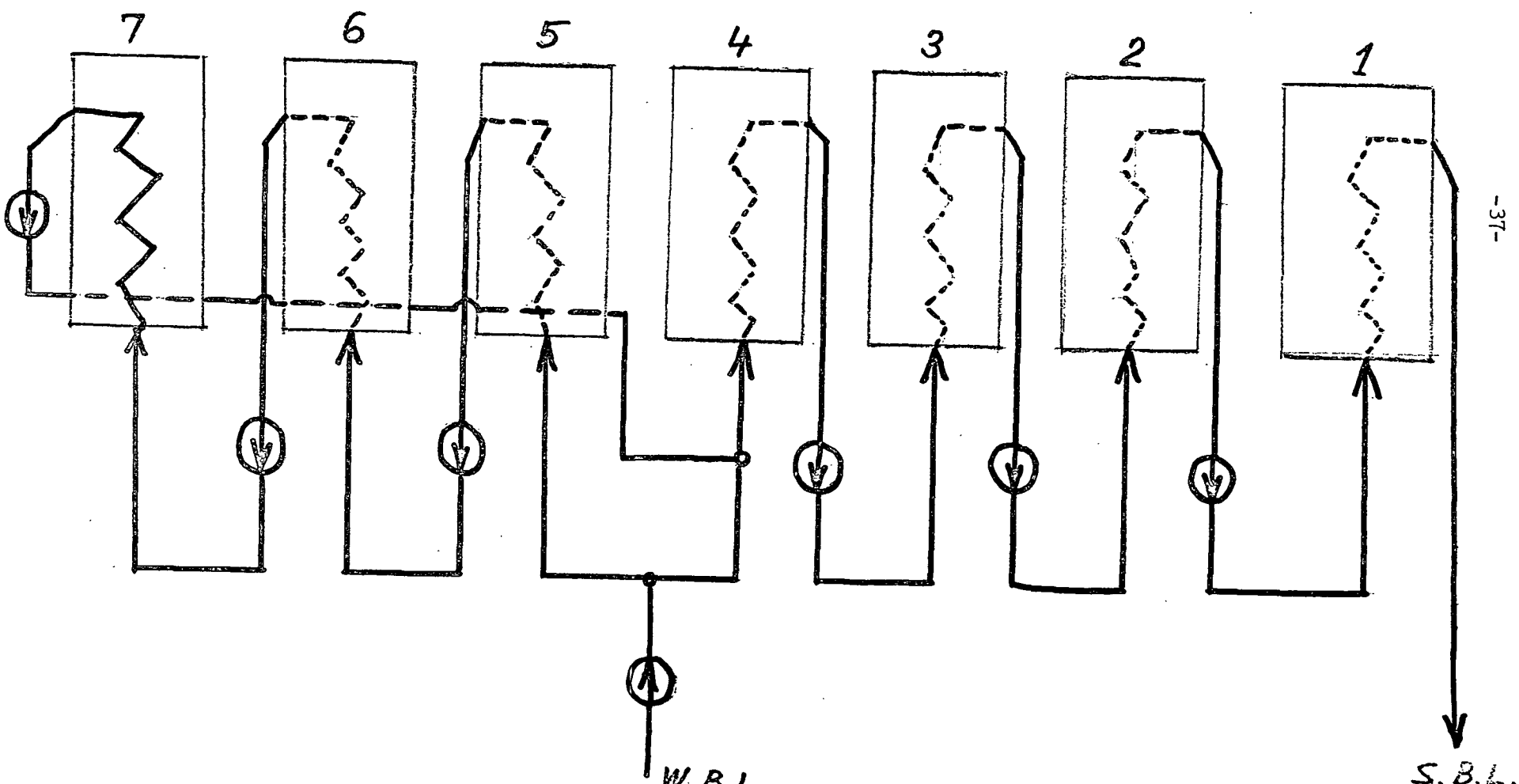
Then after differentiation a final form is:

$$\frac{dA_N}{dt} = \frac{C_P}{U_N} \cdot \frac{\Delta T\left[W_{LN} \frac{d(T_{Bi} - T_{Bi+1})}{dt} + (T_{Bi} - T_{Bi+1}) \frac{dW_{LN}}{dt}\right] - (T_{Bi} - T_{Bi+1}) W_{LN} \frac{d(\Delta T)}{dt}}{(T_S - T_{Bi})^2}$$

and a boiling zone heat transfer area is:

$$A_B = A - A_N$$

BLACK LIQUOR FLOW DIAGRAM



2. ADDITIONAL DATA PREDICTION AND EVALUATION:

Before the actual dynamic model is formed, there is a need for certain additional data. These are derived from basic data and conditions. Since it might be necessary to repeat the computations several times before one arrives at desired steady state conditions, a digital program is being used for this purpose, i. e. to enable easier changes and predictions of parameters in dynamic model equations. The program has 24 basic data inputs per each effect and computes 29 intermediate parameter values and performance errors per each effect.

Basic data set (variables are listed in the functional groups, for actual nomenclature listing see p. 28)

U_{ST} ----------Steam to liquor heat transfer coefficient.

W_C ----------Condensate flow.

W_{ST} ----------Steam Flow.

H_{ST} ----------Steam enthalpy.

H_C ----------Condensate enthalpy.

T_{STM}----------Steam temperature measured.

A_B ----------Boiling section tubes heating surface.

T_{TB}----------Boiling section tubes temperature.

A_N ----------Non-boiling section heating surface.

T_{TN} ----------Non-boiling section tubes temperature.

D ----------Tubes diameter.

F_L ----------Tubes length.

BPR ----------Boiling point rise.

W_{LN} ----------Black liquor flow in.

X_{SN} ----------Black liquor solids percent in.

X_{SO} ----------Black liquor solids percent out.

W_{LO} ----------Black liquor flow out.

R_{OL} ----------Black liquor density.

H_{LW} ----------Liquid enthalpy.

Q_{TM} ----------Total heat flow computed from previous heat balance.

D_{TM} ----------Temperature difference computed from heat balance.

BP ----------Water boiling point.

U_N ----------Heat transfer coefficient in non-boiling section.

EQUATIONS PARAMETERS AND PREDICTION ERRORS ARE AS FOLLOWS:

T_{STC}----------Steam temperature predicted.

Q_L ----------Heat loss.

E_{TS} ----------Steam temperature prediction error.

Q_{SN} ----------Heat flow from steam to non-boiling section tubes.

Q_{SB} ----------Heat flow from steam to boiling section tubes.

U_N ----------Heat transfer coefficient steam to liquor in non-boiling section.

Q_N ----------Heat flow, steam to liquor in non-boiling section.

V_N----------Non-boiling section volume.

W_{SN}----------Flow of solids in.

W_{SO}----------Flow of solids out.

X_{WN} ----------
X_{WO}----------} ratios of lbs. water to lbs. Solids in and out.

E_{WS} ----------Solids flow in and out error.

E_{VNRO} ----------Water balance.

E_{NERG} ----------Energy balance.

H_{LN} ----------
H_L ----------} Liquor input and output Enthalpies.

C_{PL} ----------Black liquor specific heat.

Q_B ----------Heat flow to liquor in boiling section.

E_{VPR} ----------Evaporation rate.

E_{RENR} ----------Energy balance error.

T_B ----------Liquor temperature.

Q_{TOT} ----------Total heat flow steam to liquor.

ERQT ----------Total heat flow error- difference between Q computed by predictor program and Q computed by heat balance program.

ERDT ----------Temperature difference error.

U_{TOT} ----------Over all heat transfer coefficient steam to liquor.

E_{RUT} ----------Heat transfer coefficient error.

Using "Heat Balance Results" from page 16

a following table containing a basic data set can be compiled.

Variable	Units	No. 1	No. 2	No. 3	No. 4	No. 5	No. 6	No. 7
Q_{TOT}	BTU/HRx10^6	64	56	60	86	62	33	59
U_{TOT}	BTU/hr°Fft^2	196	326	225	326	184	70	117
T_{ST}	°F	293	259	237	215	203	179	147
DT	°F	23	12	16	17	20	28	26
T_B	°F	270	247	221	198	183	151	121
X_{SN}	% Sol.	36	30	25.5	21	15.7	18	19.5
X_{SO}	% Sol.	47	36	30	25.5	18	19.5	22.8
W_{LI}	LBS/HRx10^3	293	352	414	480	450	430	396
W_{LO}	LBS/HRx10^3	224	293	352	414	430	396	339
BPR	°F	11	10	6	5	4	4	6
A	SQFT x 10^2	143	143	167	155	169	169	194

HEAT BALANCE RESULTS

VARIABLE	UNITS	NO.1	NO.2	NO.3	NO.4	NO.5	NO.6	NO.7	FLASH TANK
STEAM FEED	LBS/HR X 10**3	88.							
STEAM TEMP.	DEGREE F	293.	259.	237.	215.	203.	179.	147.	
TEMP.DIFFERENCE	DEGREE F	23.	12.	16.	17.	20.	28.	26.	
LIQUOR TEMP.	DEGREE F	270.	247.	221.	198.	183.	151.	121.	212.
B.P.R.	DEGREE F	11.	10.	6.	5.	4.	4.	6.	
SATUR.VAP.TEMP.	DEGREE F	259.	237.	215.	203.	179.	147.	115.	
VAPOR PRESSURE	PSIA	35.	24.	16.	12.	7.	4.	2.	
LATENT HEAT VAP	BTU/LB	940.	954.	967.	975.	990.	1010.	1029.	
DISCHARGE X1000	LBS/HR	224.	293.	352.	414.	430.	396.	339.	220.
SOLIDS IN	PERCENTS	36.0	30.0	25.5	21.0	15.7	18.0	19.5	47.0
SOLIDS OUT	PERCENTS	47.0	36.0	30.0	25.5	18.0	19.5	22.8	48.0
HT.TRANSF.COEFF	BTU/HR*DGF*SQFT	196.	326.	225.	~~798.~~ 326	184.	70.	117.	
HEATING SURFACE	SQFT X 10**2	143.	143.	167.	155.	169.	169.	194.	
Q HEAT X10**6	BTU/HR	64.	56.	60.	86.	62.	33.	59.	
STEAM ECONOMY	LBSWATER/LBSTM.	4.96							
W.B.L. TEMP.	DEGREE F					194.			
SPLIT FEED	PERCENTS				25.	75.			

DATA SAMPLED DATE 6 18 70

JOB NUMBER IS 1

HEAT BALANCE RESULTS

VARIABLE	UNITS	NO.1	NO.2	NO.3	NO.4	NO.5	NO.6	NO.7	FLASH TANK
STEAM FEED	LBS/HR X 10**3	91.							
STEAM TEMP.	DEGREE F	287.	252.	224.	0.	204.	182.	150.	
TEMP.DIFFERENCE	DEGREE F	24.	18.	14.	0.	18.	28.	19.	
LIQUOR TEMP.	DEGREE F	263.	234.	210.	0.	186.	154.	131.	222.
B.P.R.	DEGREE F	11.	10.	6.	0.	4.	4.	6.	
SATUR.VAP.TEMP.	DEGREE F	252.	224.	204.	0.	182.	150.	125.	
VAPOR PRESSURE	PSIA	31.	19.	13.	0.	8.	4.	2.	
LATENT HEAT VAP	BTU/LB	944.	962.	975.	0.	990.	1008.	1022.	
DISCHARGE X1000	LBS/HR	182.	246.	329.	0.	485.	420.	388.	177.
SOLIDS IN	PERCENTS	354.0	265.0	225.0	0.0	161.0	180.0	208.0	480.0
SOLIDS OUT	PERCENTS	480.0	354.0	265.0	0.0	180.0	208.0	225.0	493.0
HT.TRANSF.COEFF	BTU/HR*DGF*SQFT	178.	309.	244.	0.	186.	139.	88.	
HEATING SURFACE	SQFT X 10**2	143.	143.	167.	155.	169.	169.	194.	
Q HEAT X10**6	BTU/HR.	61.	79.	57.	0.	56.	65.	32.	
STEAM ECONOMY	LBSWATER/LBSSTM.	4.01							
W.B.L. TEMP.	DEGREE F					191.			
SPLIT FEED	PERCENTS				0.	100.			

DATA SAMPLED DATE 7 9 70

JOB NUMBER IS 2

NOTE: Due to a key-punching error decimal point for "Solids In" and "Solids Out" shifted one space right: example -- 354.0 should be 35.4%.

HEAT BALANCE RESULTS

VARIABLE	UNITS	NO.1	NO.2	NO.3	NO.4	NO.5	NO.6	NO.7
STEAM FEED	LBS/HR	96.						
STEAM TEMP.	DEGREE F	282.	252.	228.	0.	207.	183.	155.
TEMP.DIFFERENCE	DEGREE F	20.	16.	15.	0.	21.	24.	28.
LIQUOR TEMP.	DEGREE F	262.	236.	213.	0.	186.	159.	127.
B.P.R.	DEGREE F	10.	8.	6.	0.	3.	4.	4.
SATUR.VAP.TEMP.	DEGREE F	252.	228.	207.	0.	183.	155.	123.
VAPOR PRESSURE	PSIA	31.	20.	13.	0.	8.	4.	2.
LATENT HEAT VAP	BTU/LB	944.	960.	973.	0.	988.	1005.	1024.
DISCHARGE X1000	LBS/HR	213.	307.	373.	0.	580.	497.	435.
SOLIDS IN	PERCENTS	34.	28.	24.	0.	17.	18.	21.
SOLIDS OUT	PERCENTS	49.	34.	28.	0.	18.	21.	24.
HT.TRANSF.COEFF.	BTU/HR*DGF*SQFT	310.	276.	241.	0.	95.	205.	117.
HEATING SURFACE	SQFT X 10**2	143.	143.	167.	155.	169.	169.	194.
Q HEAT X10**6	BTU/HR	88.	63.	60.	0.	33.	83.	63.
STEAM ECONOMY	LBSWATER/LBSSTM.	4.18						
W.B.L. TEMP.	DEGREE F					197.		
SPLIT FEED	PERCENTS				0.	100.		

DATA SAMPLED 7/17/70
JOB NUMBER IS 3

HEAT BALANCE RESULTS

VARIABLE	UNITS	NO.1	NO.2	NO.3	NO.4	NO.5	NO.6	NO.7
STEAM FEED	LBS/HR	100.						
STEAM TEMP.	DEGREE F	292.	256.	234.	0.	214.	188.	157.
TEMP.DIFFERENCE	DEGREE F	25.	15.	15.	0.	24.	28.	20.
LIQUOR TEMP.	DEGREE F	267.	241.	219.	0.	190.	160.	137.
B.P.R.	DEGREE F	11.	7.	5.	0.	2.	3.	4.
SATUR.VAP.TEMP.	DEGREE F	256.	234.	214.	0.	188.	157.	133.
VAPOR PRESSURE	PSIA	34.	22.	15.	0.	9.	4.	2.
LATENT HEAT VAP	BTU/LB	941.	956.	969.	0.	985.	1004.	1018.
DISCHARGE X1000	LBS/HR	210.	315.	404.	0.	594.	532.	481.
SOLIDS IN	PERCENTS	32.	25.	21.	0.	15.	17.	19.
SOLIDS OUT	PERCENTS	48.	32.	25.	0.	17.	19.	21.
HT.TRANSF.COEFF	BTU/HR*DGF*SQFT	277.	394.	297.	0.	192.	132.	132.
HEATING SURFACE	SQFT X 10**2	143.	143.	167.	155.	169.	169.	194.
Q HEAT X10**6	BTU/HR	99.	84.	74.	0.	78.	62.	51.
STEAM ECONOMY	LBSWATER/LBSSTM.	4.63						
W.B.L. TEMP.	DEGREE F					196.		
SPLIT FEED	PERCENTS				0.	100.		

DATA SAMPLED DATE 8 10 70

~~DATA SAMPLED DATE 70~~

JOB NUMBER IS 4

HEAT BALANCE RESULTS

VARIABLE	UNITS	NO.1	NO.2	NO.3	NO.4	NO.5	NO.6	NO.7	FLASH TANK
STEAM FEED	LBS/HR X 10**3	100.							
STEAM TEMP.	DEGREE F	292.	258.	232.	0.	212.	183.	153.	
TEMP.DIFFERENCE	DEGREE F	23.	18.	15.	0.	26.	26.	23.	
LIQUOR TEMP.	DEGREE F	269.	240.	217.	0.	186.	157.	130.	223.
B.P.R.	DEGREE F	11.	8.	5.	0.	3.	4.	5.	
SATUR.VAP.TEMP.	DEGREE F	258.	232.	212.	0.	183.	153.	125.	
VAPOR PRESSURE	PSIA	34.	21.	14.	0.	8.	4.	2.	
LATENT HEAT VAP	BTU/LB	940.	958.	970.	0.	988.	1006.	1023.	
DISCHARGE X1000	LBS/HR	208.	311.	386.	0.	554.	487.	443.	203.
SOLIDS IN	PERCENTS	321.0	258.0	225.0	0.0	156.0	180.0	205.0	480.0
SOLIDS OUT	PERCENTS	480.0	321.0	258.0	0.0	180.0	205.0	225.0	491.0
HT.TRANSF.COEFF	BTU/HR*DGF*SQFT	294.	282.	219.	0.	191.	154.	99.	
HEATING SURFACE	SQFT X 10**2	143.	143.	167.	155.	169.	169.	194.	
Q HEAT X10**6	BTU/HR	96.	72.	55.	0.	84.	68.	44.	
STEAM ECONOMY	LBSWATER/LBSSTM.	4.36							
W.B.L. TEMP.	DEGREE F					191.			
SPLIT FEED	PERCENTS				0.	100.			

DATA SAMPLED DATE 8 18 70
JOB NUMBER IS 5

HEAT BALANCE RESULTS

VARIABLE	UNITS	NO.1	NO.2	NO.3	NO.4	NO.5	NO.6	NO.7	FLASH TANK
STEAM FEED	LBS/HR X 10**3	100.							
STEAM TEMP.	DEGREE F	288.	254.	230.	0.	212.	184.	155.	
TEMP.DIFFERENCE	DEGREE F	24.	16.	13.	0.	25.	25.	25.	
LIQUOR TEMP.	DEGREE F	264.	238.	217.	0.	187.	159.	130.	2220.
B.P.R.	DEGREE F	10.	8.	5.	0.	3.	4.	5.	
SATUR.VAP.TEMP.	DEGREE F	254.	230.	212.	0.	184.	155.	125.	
VAPOR PRESSURE	PSIA	32.	21.	14.	0.	8.	4.	2.	
LATENT HEAT VAP	BTU/LB	942.	958.	970.	0.	987.	1005.	1022.	
DISCHARGE X1000	LBS/HR	218.	313.	389.	0.	566.	500.	445.	209.
SOLIDS IN	PERCENTS	327.0	263.0	230.0	0.0	161.0	181.0	205.0	470.0
SOLIDS OUT	PERCENTS	470.0	327.0	263.0	0.0	181.0	205.0	230.0	489.0
HT.TRANSF.COEFF	BTU/HR*DGF*SQFT	261.	319.	249.	0.	164.	157.	114.	
HEATING SURFACE	SQFT X 10**2	143.	143.	167.	155.	169.	169.	194.	
Q HEAT X10**6	BTU/HR	89.	73.	54.	0.	69.	66.	55.	
STEAM ECONOMY	LBSWATER/LBSSTM.	4.27							
W.B.L. TEMP.	DEGREE F					194.			
SPLIT FEED	PERCENTS				0.	100.			

DATA SAMPLED DATE 8 25 70

JOB NUMBER IS 6

HEAT BALANCE RESULTS

VARIABLE	UNITS	NO.1	NO.2	NO.3	NO.4	NO.5	NO.6	NO.7	FLASH TANK
STEAM FEED	LBS/HR X 10**3	92.							
STEAM TEMP.	DEGREE F	282.	250.	228.	0.	210.	182.	155.	
TEMP.DIFFERENCE	DEGREE F	22.	14.	13.	0.	25.	23.	25.	
LIQUOR TEMP.	DEGREE F	260.	236.	215.	0.	185.	159.	130.	219.
B.P.R.	DEGREE F	10.	8.	5.	0.	3.	4.	5.	
SATUR.VAP.TEMP.	DEGREE F	250.	228.	210.	0.	182.	155.	125.	
VAPOR PRESSURE	PSIA	30.	20.	14.	0.	8.	4.	2.	
LATENT HEAT VAP	BTU/LB	945.	960.	971.	0.	989.	1005.	1022.	
DISCHARGE X1000	LBS/HR	207.	300.	354.	0.	529.	459.	403.	203.
SOLIDS IN	PERCENTS	328.0	278.0	244.0	0.0	166.0	186.0	214.0	474.0
SOLIDS OUT	PERCENTS	474.0	328.0	278.0	0.0	186.0	214.0	244.0	484.0
HT.TRANSF.COEFF	BTU/HR*DGF*SQFT	277.	258.	220.	0.	149.	179.	119.	
HEATING SURFACE	SQFT X 10**2	143.	143.	167.	155.	169.	169.	194.	
Q HEAT X10**6	BTU/HR	87.	51.	47.	0.	63.	69.	57.	
STEAM ECONOMY	LBSWATER/LBSSTM.	4.23							
W.B.L. TEMP.	DEGREE F					197.			
SPLIT FEED	PERCENTS				0.	100.			

DATA SAMPLED DATE 9 4 70

JOB NUMBER IS 7

```
      PROGRAM  SSDTEV
      DIMENSION UST(7),WC(7),WST(7),HST(7),HC(7),TSTM(7),AB(7),TTB(7)
      DIMENSION AN(7),TTN(7),BPR(7),WLN(7),XSN(7),XSO(7),WLO(7),ROL(7)
      DIMENSION HW(7),QTM(7),DTM(7),UTM(7),X(7),TSTC(7),QL(7),ETS(7)
        DIMENSION QN(7),WSN(7),WSO(7),XWN(7),XWO(7),EWS(7),EVNRO(7)
      DIMENSION ENERG(7),CPLN(7),HL(7),HLN(7),QB(7),EVPR(7),ENRG(7)
      DIMENSION ENRG1(7),ERENR(7),TB(7),QTOT(7),ERQT(7),AT(7),DT(7)
      DIMENSION ERDT(7),UTOT(7),ERUT(7),UN(7) ,CPL(7),VN(7),PT(7)
      DIMENSION QSR(7),QSN(7),BP(7)
      DIMENSION HLW(7)
      READ(1,100)(UST(I),I=1,7)
      READ(1,100)(WC(I),I=1,7)
      READ(1,100)(WST(I),I=1,7)
      READ(1,100)(HST(I),I=1,7)
      READ(1,100)(HC(I),I=1,7)
      READ(1,100)(TSTM(I),I=1,7)
      READ(1,100)(AB(I),I=1,7)
      READ(1,100)(TTB(I),I=1,7)
      READ(1,100)(AN(I),I=1,7)
      READ(1,100)(TTN(I),I=1,7)
      READ(1,101)D,EL
  101 FORMAT(2F10.3)
      READ(1,100)(BPR(I),I=1,7)
      READ(1,100)(WLN(I),I=1,7)
      READ(1,100)(XSN(I),I=1,7)
      READ(1,100)(XSO(I),I=1,7)
      READ(1,100)(WLO(I),I=1,7)
      READ(1,100)(ROL(I),I=1,7)
      READ(1,100)(HLW(I),I=1,7)
      READ(1,100)(HW(I),I=1,7)
      READ(1,100)(QTM(I),I=1,7)
      READ(1,100)(DTM(I),I=1,7)
      READ(1,100)(UTM(I),I=1,7)
      READ(1,100)(BP(I),I=1,7)
      READ(1,100)(PT(I),I=1,7)
```

```
STEAM CHEST

DO 301 I=1,7
X(I)=WST(I)*HST(I)-WC(I)*HC(I)
TSTC(I)=(WST(I)*(HST(I)+32.)+UST(I)*(AB(I)*TTB(I)+AN(I)*TTN(I)))
1/(WST(I)+UST(I)*(AB(I)+AN(I)))
QL(I)=-TSTM(I)*(WST(I)+UST(I)*(AB(I)+AN(I)))+WST(I)*(HST(I)+32.)+
1UST(I)*(AB(I)*TTB(I)+AN(I)*TTN(I))
PTS(I)=TSTC(I)-TSTM(I)
QSB(I)=UST(I)*AB(I)*(TSTC(I)-TTB(I))
QSN(I)=UST(I)*AN(I)*(TSTC(I)-TTN(I))

NON-BOILING SECTION,TUBES

QN(I)=UN(I)*AN(I)*(TST(I)-TN(I))

NON-BOIL.SECTION - LIQUOR

VN(I)=3.14*D**2.*PT(I)/4.

WSN(I)=WLN(I)*XSN(I)
WSO(I)=WLO(I)*XSO(I)
XWN(I)=1.-XSN(I)
XWO(I)=1.-XSO(I)
EWS(I)=WSN(I)-WSO(I)
EVNRO(I)=(WSN(I)*XWN(I))-(WSO(I)*XWO(I))
ENERG(I)=WSN(I)*(1.+XWN(I))*HLN(I)-WSO(I)*(1.+XWO(I))*HL(I)+QN(I)
HLN(I)=HLW(I)+CPLN(I)*TTN(I)
CPLN(I)=.990+8.0*.00001*TTN(I)-(.6390-6.4*.0001*TTN(I))*XSN(I)
HL(I)=HLW(I)+CPL(I)*TTN(I)
```

```
C
C       THE BOILING SECTION
C
C
        QB(I)=UST(I)*AB(I)*(TSTC(I)-TTB(I))
C
C       THE LIQUOR IN THE BOILING SECTION
C
        EVPR(I)=WSN(I)*(XWN(I)-XWO(I))
        ENRG(I)=WSN(I)*(1.+XWN(I))*HLN(I)+QB(I)
        ENRG1(I)=WSN(I)*(1.+XWO(I))*HL(I)+EVPR(I)*HW(I)
        ERENR(I)=ENRG(I)-ENRG1(I)
        TB(I)=BP(I)+BPR(I)
C
C
C       OVER-ALL DATA
C
C
        QTOT(I)=QN(I)+QB(I)
        ERQT(I)=QTOT(I)-QTM(I)
        AT(I)=AN(I)+AB(I)
        DT(I)=TSTC(I)-TTB(I)
        ERDT(I)=DT(I)-DTM(I)
        UTOT(I)=QTOT(I)/(AT(I)*DT(I))
        EPUT(I)=UTOT(I)-UTM(I)
  301   CONTINUE
  100   FORMAT(7F10.3 )
        WRITE(3,200)
  200   FORMAT(1H1 )
        WRITE(3,201)
  201   FORMAT(38HRESULTING DATA AND STEADY STATE VALUES)
        WRITE(3,202)
  202   FORMAT(11HSTEAM CHEST)
        WRITE(3,203)
  203   FORMAT(25HSTEAM TEMP. COMPUTED TSTC)
        WRITE(3,204)(TSTC(I),I=1,7)
  204   FORMAT(7F5.1)
        WRITE(3,205)
  205   FORMAT(16HTHE HEAT LOSS QL)
        WRITE(3,206)(QL(I),I=1,7)
  206   FORMAT(7F10.0)
        WRITE(3,207)
  207   FORMAT(32HCOMP. MINUS MEASURED STEAM TEMP.)
        WRITE(3,204)(ETS(I),I=1,7)
        WRITE(3,209)
  209   FORMAT(47HHEAT FLOWS TO THE TUBE WALLS NON AND BOIL.SECT.)
        WRITE(3,206)(QSN(I),I=1,7)
```

```
      WRITE(3,206)(QSB(I),I=1,7)
      WRITE(3,200)

      WRITE(3,210)
  210 FORMAT(23HNON-BOILING SECT. TUBES)
      WRITE(3,211)
  211 FORMAT(21HHEAT TRANSF. COEF. UN)
      WRITE(3,204)(UN(I),I=1,7)
      WRITE(3,212)
  212 FORMAT(40HHEAT FLOW STEAM CHEST LIQUOR N.B.SECT.QN)
      WRITE(3,206)(QN(I),I=1,7)
      WRITE(3,213  )
  213 FORMAT(23HNON BOIL.SECTION LIQUOR)
      WRITE(3,235)
  235 FORMAT(32HNON-BOILING SECTION VOLUME,VN(I))
      WRITE(3,206)(VN(I),I=1,7)
      WRITE(3,214)
  214 FORMAT(17HFLOW OF SOLIDS IN)
      WRITE(3,206)(WSN(I),I=1,7)
      WRITE(3,215)
  215  FORMAT(18HFLOW OF SOLIDS OUT)
      WRITE(3,206)(WSO(I),I=1,7)
      WRITE(3,216)
  216 FORMAT(45HRATIOS OF LB.WATER TO LB.OF SOLIDS IN AND OUT)
      WRITE(3,217)
  217 FORMAT(28HSOLIDS FLOW IN AND OUT ERROR)
      WRITE(3,206)(EWS(I),I=1,7)
      WRITE(3,218)
  218 FORMAT(25HWATER DIFFERENCE EVNRO(I))
      WRITE(3,206)(EVNRO(I),I=1,7)
      WRITE(3,219)
  219 FORMAT(23HENERGY BALANCE ENERG(I))
      WRITE(3,206)(ENERG(I),I=1,7)
      WRITE(3,221)
  221 FORMAT(34HENTHALPIES IN AND OUT,HLN(I),HL(I))
      WRITE(3,206)(HLN(I),I=1,7)
      WRITE(3,206)(HL(I),I=1,7)
      WRITE(3,222)(CPLN(I),I=1,7)
  222 FORMAT(7F10.4)
      WRITE(3,222)(CPL(I),I=1,7)
      WRITE(3,200)
      WRITE(3,223)
  223 FORMAT(19HTHE BOILING SECTION)
      WRITE(3,224)
  224 FORMAT(38HHEAT FLOW TO LIQUOR IN BOILING SECTION)
      WRITE(3,206)QB(1),QB(2),QB(3),QB(4),QB(5),QB(6),QB(7)
```

```
      WRITE(3,225)
  225 FORMAT(33HTHE LIQUOR IN THE BOILING SECTION)
      WRITE(3,226)
  226 FORMAT(24HEVAPORATION RATE,EVPR(I))
      WRITE(3,206)(EVPR(I),I=1,7)
      WRITE(3,227)
  227 FORMAT(29HENERGY BALANCE ERROR,ERENR(I))
      WRITE(3,206)(ERENR(I),I=1,7)
      WRITE(3,228)
  228 FORMAT(25HLIQ. TEMP. COMPUTED,TB(I))
      WRITE(3,204)(TB(I),I=1,7)
      WRITE(3,200)
      WRITE(3,229)
  229  FORMAT(13HOVER ALL DATA)
      WRITE(3,230)
  230 FORMAT(30HTOTAL HEAT TRANSFERRED,QTOT(I))
      WRITE(3,231)
  231  FORMAT(73HTOTAL HEAT ERROR-DIFFERENCE BETWEEN COMPUTED QTOT AND Q
     1 FROM DATA SAMPLED)
      WRITE(3,206)(ERQT(I),I=1,7)
      WRITE(3,232)
  232 FORMAT(29HTEMP.DIFFERENCE ERROR,ERDT(I))
      WRITE(3,204)(ERDT(I),I=1,7)
      WRITE(3,233)
  233 FORMAT(33HOVER ALL HEAT TRANSF.COEF,UTOT(I))
      WRITE(3,204)(UTOT(I),I=1,7)
      WRITE(3,234)
  234 FORMAT(31HUTOT(I) ERROR FROM SAMPLED DATA)
      STOP
      END
```

THE NOMENCLATURE LISTING:

A_B ------ Outer area of tube walls of boiling section (sq. ft.)

A_N------- Outer area of tube walls of non-boiling section (sq.ft.)

A_{NM}------ Mean area of tube walls of non-boiling section (sq.ft.)

α ------ Constant (.00075 lbs/°F(cu. ft.)

β ------ Constant (-12.2×10^{-2})

BPR ----- Boiling point rise (°F)

$C_{P\ell}$ ----- Specific heat of black liquor (Btu/lb °F)

D ------ Tube diameter (ft.)

H_ℓ ------ Enthalpy of black liquor (Btu/lb.)

H_{ST}------ Enthalpy of steam (Btu/lb.)

FL ------ Tube length (ft.)

M_{TN}------ Mass of tube walls of non-boiling section (lb.)

P_A ------ Acceleration pressure (lb. force/sq.ft.)

P_F------- Friction pressure drop (lb.force/sq.ft.)

P_H------- Pressure from static head (lb. force/sq.ft.)

ΔP_{TOT}---- Total pressure drop across tubes (lb.force/sq.ft.)

Q_B------- Heat flow to liquor in boiling zone(Btu/hr.)

Q_L ------ Heat loss (Btu/hr.)

Q_{SB} ----- Heat flow from steam to boiling section (Btu/hr.)

Q_{SN} ----- Heat flow from steam to non-boiling section.

Q_{ST} ----- Heat flow from steam to tubes (Btu/hr.)

R_W ------ Evaporation rate (lb./hr.)

T_B------- Temperature of liquor in boiling section (°F)

T_N ------ Temperature of liquor in tubes of non-boiling section (°F)

T_{Bi}------ Temperature in i-th effect (°F)

T_{Bi+1} ----------Temperature in previous effect (°F)

T_{ST} ---------- Steam temperature (°F)

T_{TB} ---------- Temperature of tube walls of boiling section (°F)

T_{TN} ---------- Temperature of tube walls of non-boiling section (°F)

U_B ---------- Heat transfer coefficient steam chest-liquor in boiling section (Btu/sq.ft.°F hr.)

U_N ---------- Heat transfer coefficient steam chest - liquor in non-boiling section.

U_{ST}---------- Heat transfer coefficient steam - tube walls (Btu/sq.ft.hr.°F)

V_N ---------- Volume of liquor in non-boiling section (cu.ft.)

V_{ST} ---------- Volume of steam chest (cu. ft.)

W_C ---------- Condensate flow (lb/hr.)

W_S ---------- Flow of solids in liquor (lb./hr.)

W_{ST} ---------- Steam flow (lb./hr.)

X_W ---------- Fraction of water in black liquor to solids. (lb./lb.)

X_S ---------- Fraction of solids in black liquor to total weight (lb./lb.)

ρ_l ---------- Density of solids (lb./cu.ft.)

ρ_{ST}---------- Density of steam (lb./cu.ft.)

τ ---------- Time constant (min.)

τ_l ---------- Time constant of liquor temperature (min.)

τ_{ST} ---------- Time constant of steam temperature (min.)

τ_T ---------- Time constant of tube wall temperature (min.)

SUBSCRIPTS:

1 -----Input

0 -----Output

5. LITERATURE REFERENCES

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4. Eckert, E. R. G., "Heat and Mass Transfer," McGraw Hill, 1969.

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THE MATHEMATICAL MODELING AND CONTROL OF THE RECOVERY UNIT

BY

FRODE L. GALTUNG

December 10, 1970

TABLE OF CONTENTS

1. INTRODUCTION

The present project is directed at the problem of developing a mathematical model of and a practical digital computer control scheme for a kraft process recovery unit consisting of its furnace and related devices.

The practice of recovery unit control is presently directed more towards stability of operation than optimization of costs or returns. The major reason for this is the concern for safety. It is believed that a more advanced control system than is used at present will not only be able to achieve stable operation but also operation at the optimum operating point. As safety is important, it will be desirable to have automated emergency shut-down procedures that are initiated automatically when a hazardous situation is detected. Completely automatic control would permit the recovery unit to be removed from the rest of the process area so that it no longer would be considered a hazard to the operator. This will require that daily maintenance operations that are now performed by the operators will have to be either automated or eliminated.

The student had summer employment at the Owens-Illinois kraft pulp mill at Valdosta, Georgia, where three Combustion Engineering recovery units were operated in parallel. One of the units was only two years old, and offered an excellent opportunity of studying a furnace where most of the instrumentation was in working order. The operation of the furnaces under various conditions was studied, and some data were collected.

During the fall the control system for the combustion air has been analyzed, and the possibility for completely automatic

combustion air control has been discussed. A mechanism for the dynamics and the chemical reactions in the char bed is proposed.

Only the Combustion Engineering system for recovery units will be discussed specifically in this project, and particularly data from Valdosta will be used wherever possible. It is believed that in spite of a number of differences between the Combustion Engineering and the Babcock & Wilcox recovery units, their basic principles of operation are the same. Most of the conclusions regarding the operation and control of the Combustion Engineering units should apply with possibly only slight modifications also to the Babcock & Wilcox units.

The student is indebted to Mr. F.A. Rummel, Dr. N.C.S. Chari and the entire Valdosta Mill staff for their kind cooperation in this project.

2. COMBUSTION AIR CONTROL

2.1 Current Control System and Operating Practice

A modern combustion air control system is shown schematically in Figure 1. The black liquor flow set point is relayed to the set points of the primary and the total air flow controllers. A change in the set point to the black liquor flow controller will automatically cause a proportional change in the set points of the two air flow controllers. An internal draft control system is independent of the combustion air control system. Furnace draft is controlled by varying the speed of the I.D. fan.

Instrumentation in addition to what is required for the automatic control loops usually includes oxygen and combustibles analyzers for analysis of the flue gases, and pressure indi-

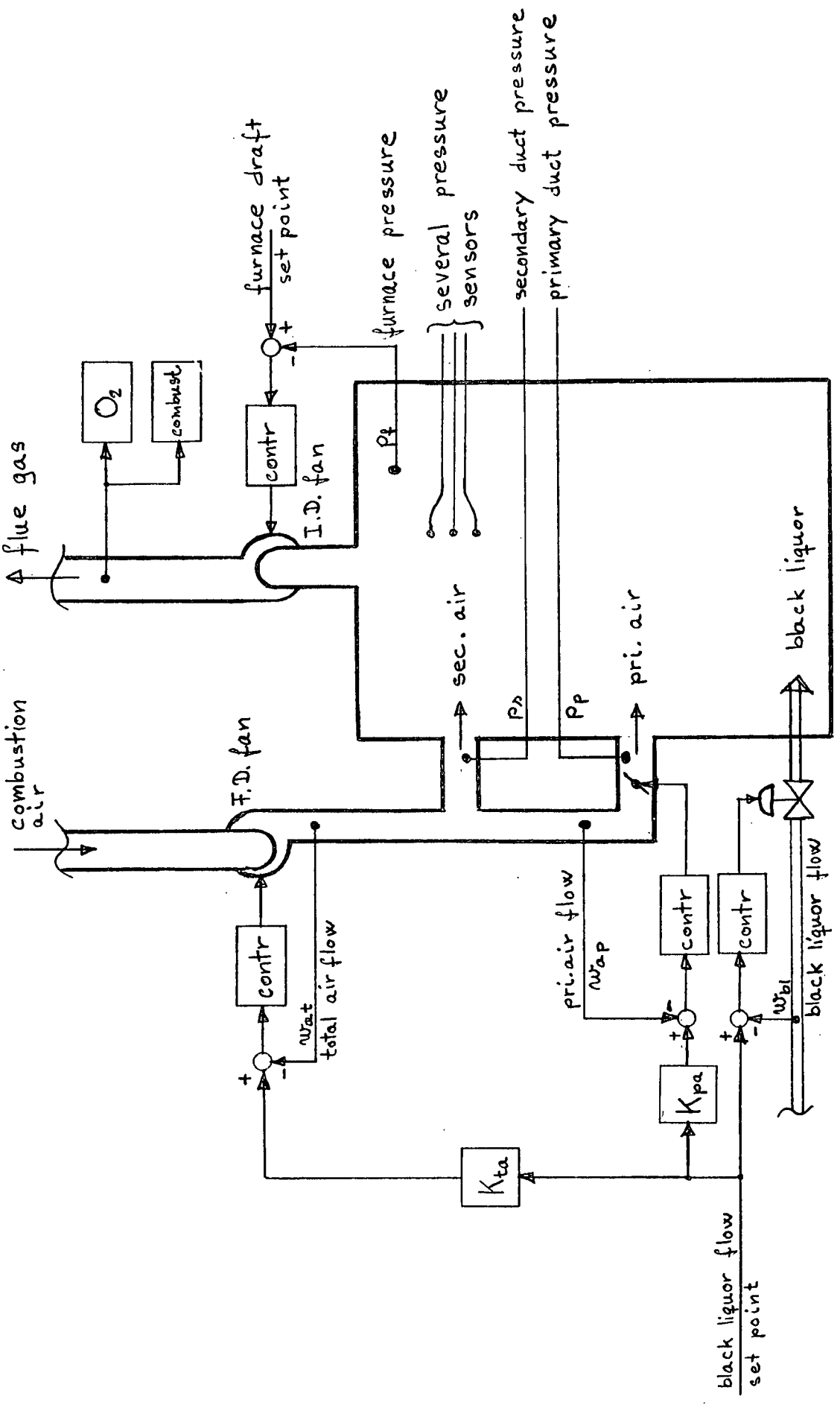

Figure 1

Combustion air control system

cators for monitoring of the pressures at different locations in the furnace and in the air and flue gas ducts. These instruments are only indicating, however, and not part of the automatic control system. They may be connected to the annunciator panel so that abnormal readings cause alarms to be given off.

The present system requires an operator to perform the following duties as part of the combustion air control:

--He must adjust the coefficients K_{ta} and K_{pa} in Figure 1 whenever the composition of the black liquor changes. The most normal variations are due to changes in the concentration of total solids in the liquor. A large amount of experience is required to be able to change these coefficients properly The operator wants to maintain a constant oxygen deficiency in the primary zone in order to produce smelt of high reduction ratio. Too much primary air (too large K_{pa}) will lower the reduction ratio,-too little will give insufficient combustion and the char bed will "blacken out" due to a deficiency of heat. Unfortunately there is usually a large time delay between the moment when action is taken by the operator and when the reduction analysis from the green liquor is available. (Routine reduction tests are taken once every 24 hours.)

When K_{pa} is fixed, then variations in K_{ta} will change the flow of secondary air. Too much secondary air will decrease the efficiency of the steam generating system as air that does not participate in the combustion process will pass through the furnace and carry away parts of the heat that should have been converted into steam. Too little secondary air will result in incomplete combustion. This both decreases the effi-

ciency of the steam generation as well as sharply increasing the carbon monoxide and sulfur content of the flue gas. The operator has better guidance for the adjustment of K_{ta} than for K_{pa}. The flue gas analyzers respond almost immediately to changes in the total air flow. During normal operation the combustibles analysis should be close to a zero reading (on a 0-10% (Volume) scale) and the oxygen analysis should be approximately 3% (Volume). A deficiency in total air will result in a low oxygen and a high combustibles analysis. An excess of total air will result in a low (normal) combustibles and a too high oxygen analysis.

--The degree of turbulence in the combustion zone is also controlled by the operator. The velocity of the air entering the furnace from the air ports determines how well the air and the combustion gases will be mixed. Good mixing will yield complete combustion with a minor excess of air and this results in good steam generation efficiency. The velocity of the entering air is determined by the pressure differential between the windbox and the furnace. A constant pressure under varying flow is maintained by restricting or enlarging the passage from the windbox to the furnace. This is accomplished by manual adjustment of the windbox dampers. Normal windbox pressures may be 5" and 3" (H_2O) for the secondary and the primary windbox respectively. Unsatisfactory mixing will result in a higher-than-normal combustibles analysis and at the same time a normal -or high oxygen analysis.

--Both the primary and the secondary air ports have to be cleaned out at regular intervals. Clogging occurs gradually

as smelt floating down on the inside of the furnace walls will solidify when it is exposed to the relatively cool combustion air entering through the air ports. If the air ports are not cleaned the solid smelt will decrease the effective area of the ports, and as the air flow control system will try to maintain constant air flow, the windbox pressure will increase. The windbox pressures are monitored and displayed in the control room. Cleaning of the air ports is done manually by pushing steel rods through the air ports from the outside of the furnace and hence by breaking off solid smelt.

2.2 Proposed Computer Control System

The control scheme indicated in Figure 2 is based on the assumption that an on-line digital computer is available. Steps necessary in order to automate the actions now taken by the operator during normal furnace operation are discussed.

The main differences between this control scheme and the one in Figure 1 are:

--Primary and secondary air flows are made proportional to the total organic solids flow to the furnace -not only to the total flow of black liquor. Automatic compensation for variations in total solids content of black liquor will therefore be achieved. Known changes in the ratio between organics and inorganics in black liquor solids can be compensated for. These changes will occur if a change in the average yield from the digesters occur, -or if black liquor from different sources is being burned.

--Given the desired apparent oxygen deficiency in the primary zone and the composition and flow of black liquor, the

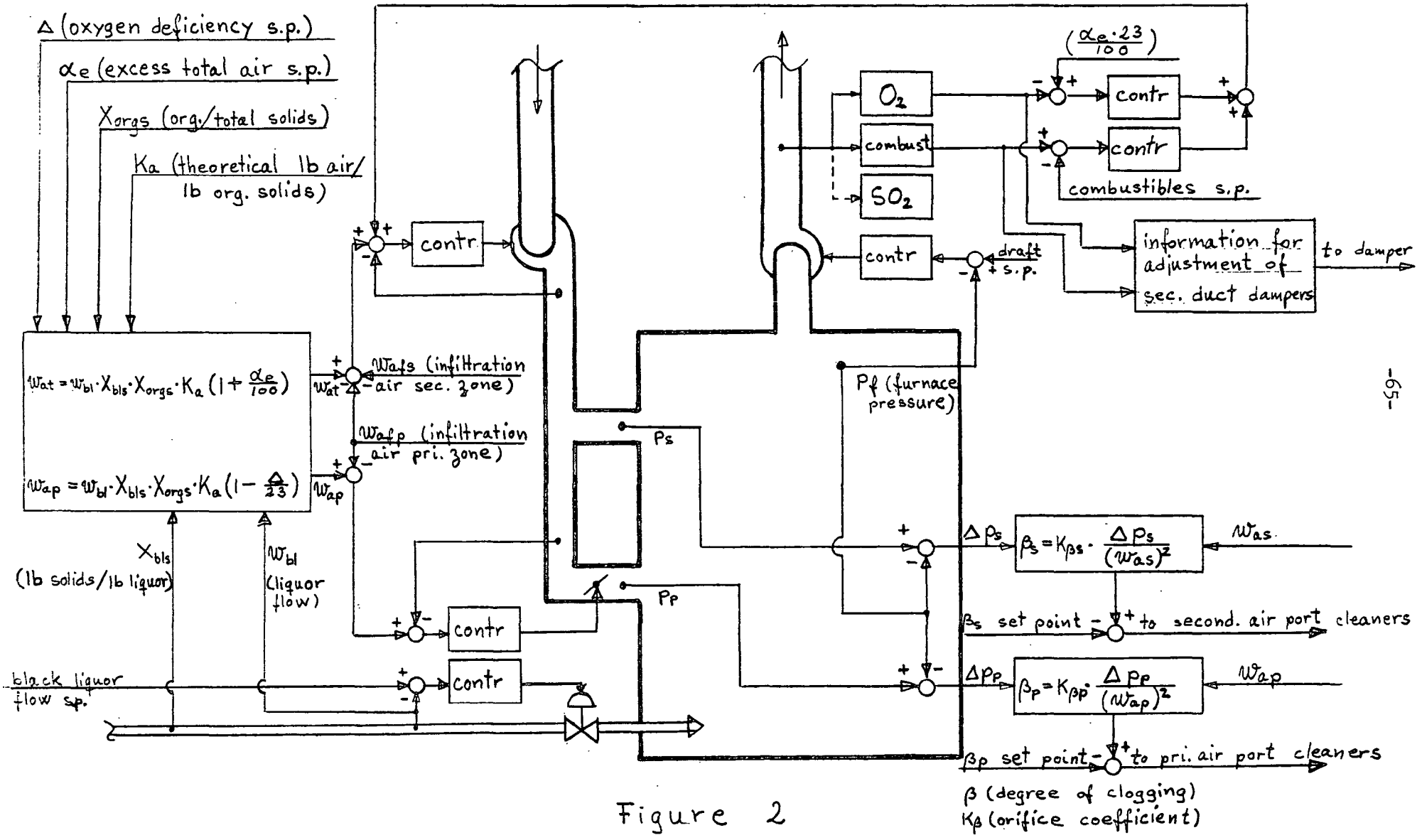

Figure 2

Computer Control of Combustion Air

flow of primary combustion air is computed. Only the concentration of total solids and the ratio between organics and inorganics in the solids are considered as variables in the composition of black liquor. The primary air control is feed-forward as no practical fast responding feedback of the reduction state of the smelt is expected to be available.

--Given the flow and composition of black liquor and the desired amount of excess air, the flow of total air is computed. Primary air and estimated infiltration air is subtracted and the difference is sent to the secondary air ports. Feedback signals from the oxygen and combustibles analyzers are used to correct the feed-forward computation of total combustion air. (A possible feedback signal from an SO_2-analyzer is indicated in case SO_2 content of the flue gas is not adequately correlated to the oxygen and combustibles content.) Automatic adjustment of the windbox dampers is desirable, and hence some motor arrangement will be required.

--The state of clogging of the air ports is measured by comparing the differential pressure across the air ports to the square of the air flow through the port. The ratio of the differential pressure to the square of the flow should be independent of the flow as long as the state of clogging is unchanged. As the clogging increases, the differential pressure will increase for constant flow. Automatic cleaning may be performed either regularly at constant intervals, -or when a certain degree of restriction of the air ports is detected. The cleaning may be performed by mechanical means, or possibly by a system of steam nozzles in the air ports.

The instrumentation includes:

--Flowmeter for black liquor

--Instrument for measurements of concentration of total solids in the black liquor

--Pressure sensors in furnace and in windboxes

--Flowmeters for total air and primary air

--Analyzers for oxygen and combustibles in the flue gas.

All these instruments are included in the control system in Figure 1, but there no direct use is made of the available information. The ratio between organics and inorganics in black liquor solids must be computed on the basis of the yield from the digesters.

--An instrument for analyzing the SO_2, -and possibly also the reduced sulfur compounds in the flue gases is indicated in Figure 2. The necessity for such an instrument depends on the correlation between the sulfur content of the flue gases with the information available from the oxygen and combustibles analyzers [1], [2]. The sulfur loss due to the sulfur content of the flue gases may be large.

3. ASSUMPTIONS FOR THE MATHEMATICAL MODEL

3.1 Some Observations regarding the Combustion Mechanism

The operation of a Combustion Engineering recovery furnace at Owens-Illinois Inc. Valdosta Mill, Valdosta, Georgia was studied rather closely during the summer 1970. As the opportunity to perform controlled experiments with the furnace was limited, conclusions regarding the process mechanisms have to be based on observations made during normal operation and on the literature.

One quickly observes the surprisingly strong effect a minor change in the temperature of the black liquor has on the furnace operation. The effect on the drop size distribution of the spray is most apparent. A finely divided spray will to a large extent be carried by the combustion gases. A too coarse spray may quickly cool down the char bed surface and cause a "black out". It is believed that the cooling effect of the coarse spray is mostly due to a larger amount of water reaching the char surface. However, as the total amount of water entering the furnace is unchanged, it seems that the water flow to the char surface has a controlling effect on the combustion rate. "Blackening" of the furnace can occur very quickly, -in the order of seconds. It is therefore possible that only a surface layer of the char bed is taking part in the initial temperature drop. This is made even more probable considering the porosity of the char bed, and that heat from the bulk of the bed has to be transferred by conduction, which is a relatively slow process.

Two basically different models of the combustion process appear to be possible:

-It may be assumed that the char bed burns as a bed of charcoal. This means that most of the energy is released on the surface of the solid particles, and that very little volatile material is given off. In order to explain the strong cooling effect of the water, one may assume that wet liquor physically prevents the primary air from coming in sufficient contact with the char particles to sustain combustion. Very little flame action is taking place.

-It may on the other hand also be assumed that char particles when heated give off large quantities of volatile material. The volatiles will burn in the gas zone above the char bed surface. Relatively little heat will be generated by the combustion of solid material. Heat is transferred to the char surface from the gas zone by radiation. The strong cooling effect of water may now be explained by the effect of temperature on the rate of distillation of volatiles from the surface char particles, and the resulting positive feedback due to a corresponding change in the radiated heat from the burning gases. Much flame action is taking place.

The actual combustion mechanism in the furnace will of course consist of both of the above mentioned, but in order to simplify the model it is desirable to use only the one that most accurately represents the process. No definite conclusions with regard to the choice of mechanism have yet been made. Observations made this summer indicate that vigorous flame action is present in most of the primary and secondary zone, -except possibly in a narrow region directly above the surface of the char bed. (Only the lowest parts of the char bed could normally be observed as one had to look through the primary air ports.) An article by J.F. Thomas and coworkers [3] describes qualitatively the initial stages of the combustion of black liquor as distillation and cracking. A rough estimate on the order of magnitude of the heat that could be expected to be radiated from the gas to the char surface was made. As example was used the recovery furnace (#3) at Valdosta and some very simplifying assumptions. We assume unity

emissivity for the gas and unity absorption coefficient for the char bed surface. The area of the char surface was taken equal to the crossection of the furnace.

Radiant heat flow: $\dot{Q} = \sigma \cdot T^4$ [Btu/hr.sq.ft]

$\sigma = 0.173 \cdot 10^{-8}$ [Btu/hr.sq.ft.$^\circ R^4$]

Gas temperature: $T_g = 2200$ [°F]

Char surface temperature: $T_c = 1500$ [°F]

Char surface area: $A_c = 353$ [sq.ft]

Black liquor flow: $W_{bl} = 67000$ [lb/hr]

Solids concentration: $x_{bls} = 0.65$ [lb/lb]

Net flow of heat to char surface:

$\dot{Q}_c = 0.173 \cdot 10^{-8} [(2660)^4 - (1960)^4] \cdot 353 = 21.7 \cdot 10^6$ [Btu/hr]

We may compare this to the latent heat of all the water flowing into the furnace:

Water flow: $67000 \cdot 0.35 = 23.4 \cdot 10^3$ [lb/hr]

Latent heat flow: $23.4 \cdot 10^3 \cdot 970 = 22.7 \cdot 10^6$ [Btu/hr]

Under normal operating conditions only a small fraction of the water is assumed to actually reach the char bed surface. Most of the water is believed to evaporate from the spray droplets while they are falling. Under the above simplifying assumptions it seems possible that radiation from the gas to the char surface may account for a major portion of the char surface energy balance.

3.2 Basic Assumptions for a Mathematical Model of the Char Bed Energy Balance

It is desirable to arrive at a model that can be used for the prediction of char bed height, smelt composition and generation rate, and the "black out condition".

Some of the basic assumptions for a mathematical model of the energy balance (given the second of the two above mentioned mechanisms) may be taken as:

1. We assume that all of the released energy is due to the combustion of volatiles given off by the liquor particles.
2. Energy supply to the char bed surface is by radiation from the burning gases only.
3. The rate of emission of volatiles is a linear function of the temperature of the char particles. (An exponential relationship may be more accurate for a large range in temperature.)
4. The char bed is lumped into two regions for energy balance purposes, - the surface layer and the bulk. Energy balance is taken with respect to both regions.
5. A fraction of the incoming black liquor spray is immediately entrained by the combustion gases. This fraction is

$$W_{bl} \cdot \eta_1 \ [\text{lb/hr}]$$

where W_{bl} is the flow of black liquor to the furnace [lb/hr]. η_1 is a function of droplet size and gas velocity in the primary zone. This fraction is responsible for SO_2 generation. It is also responsible for Na_2CO_3 and Na_2SO_4 carryover to precipitator, cascade evaporator and slag deposit surfaces. We assume the entrained fraction of the spray to burn in suspension in the primary and the secondary zone and the products of combustion to be

completely oxidized.

6. A part of the water in the black liquor spray will evaporate in suspension. The rate of evaporation is a function of droplet size, gas velocity and gas temperature.

7. The water that hits the surface of the char bed is assumed to evaporate immediately upon contact.

The modeling of η_1 and of the water that evaporates from the droplets in the spray is not completed.

3.3 Some Observations regarding the Reduction Mechanism

Observations were attempted of the effect that some of the process variables had on the smelt reduction ratio. However, only unsatisfactorily controlled experiments could be conducted, and many questions are left unanswered.

One experiment was made in order to investigate the relationship between the primary/secondary air ratio and the resulting reduction ratio. The furnace was run for 2 hours under normal operating conditions, and a composite sample was taken from the green liquor tank. The primary air was then increased from 62.5% to 70% (on a 0-100% scale which may not be very accurate). The total air and the other variables, - except secondary air were kept constant. Composite samples were taken from the green liquor tank during the next 2 hours. Later analysis showed that the reduction ratio was 93.5% and 93.7% for the first and the second 2-hour period respectively. At least two explanations for the lack of effect are possible. It may be that even at 70% primary air (this is only the reading of the instrument, - no calibrations were made)

the oxygen deficiency is large enough to cause practically complete reduction, and hence no change is experienced. The equilibrium diagram of T.W. Bauer and R.M. Dorland [4], (pp. 100), -Figure 3, indicates a reduction ratio greater than 99% at a temperature of 1600°F (which we pick arbitrarily as an average smelt temperature in the furnace) and for an oxygen deficiency larger than approximately 7% - which corresponds to an air deficiency of approximately 35%. This is roughly the same as 65% primary air. Due to insufficient mixing the actual oxygen deficiency in the primary zone may have been larger than the theoretical for 65% primary air flow, and hence the lack of response in the reduction ratio may be explained by the equilibrium diagram, - if the diagram can be considered valid for the furnace reduction ratio.

The other possible explanation is that a reduction change would have occurred if we had waited more than 2 hours. One then has to assume a long term effect from the carbon in the bulk of the char bed on the reduction ratio. It may be assumed that the smelt generated on the char surface responds to temperature and the average oxygen deficiency in the primary zone in a manner somewhat similar to what is described by Bauer and Dorland(4). However, when the smelt floats through the bulk of the char bed, whatever carbon is in there will act as a reducing agent, and compensate for a low reduction ratio on the char surface. The effect from the bulk of the char will cease when all the carbon has been used up. This mechanism is much more complicated than the previous one. It may, however, explain why experience seems to show that a reasonably high char bed aids in producing smelt of high reduction ratio.

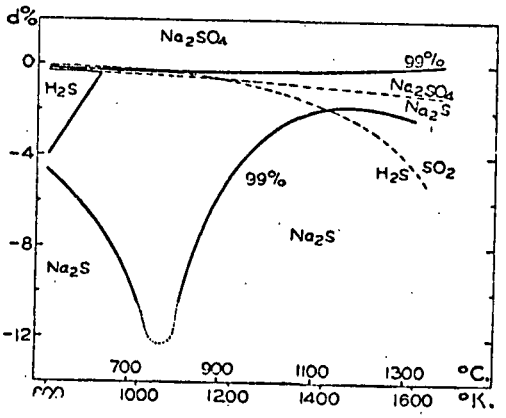

Figure 3

Equilibrium composition of smelt, [4], (pp. 100)

One experiment was made in order to investigate the relationship between char bed height and the smelt reduction ratio. The furnace was run for 2 hours under normal operating conditions and the green liquor was sampled. In order to burn down the char bed, the liquor temperature was raised. The char bed disappeared after one hour, and operation continued for another 1 3/4 hour while the green liquor was sampled. (The char bed was reduced rather slowly in order to keep the amount of carryover from the sprays to a minimum.) Analysis showed that the reduction ratio was 94.6% and 92.3% for the first and the second 2-hour period respectively. The primary air flow indicator showed 70% during the experiment. The experiment shows that reasonably high reduction ratio can be achieved without a char bed. Unfortunately prolonged operation without a char bed in the furnace is an unsafe condition, and further experimenting was not recommended. It would have been interesting to study the effect of changes in the primary/secondary air ratio on the reduction ratio when the char bed was absent.

3.4 Basic Assumptions for a Mathematical Model of the Reduction Mechanism

From the two experiments it seems that reasonably good reduction is achieved, no matter what the operator does. Experience and the routine laboratory reports prove this not to be true, however.

We know that given sufficient oxygen black liquor will burn completely and we get zero reduction ratio. If the equilibrium diagrams of Bauer and Dorland can be accepted as valid

for the average reduction ratio in the smelt, then a simple mechanization of the diagram (4 , pp. 100), - Figure 3, may be used as a first approximation to a mathematical model of the reduction mechanism. The diagram has two inputs:

-The reaction temperature may be taken as the temperature of the smelt in the furnace, which may be set equal to the temperature of the bulk of the char bed.

-The oxygen deficiency may be computed as the average theoretical oxygen deficiency in the primary zone.

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THE MATHEMATICAL MODEL OF LIME KILN

BY

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1. INTRODUCTION

In the kraft pulping process, wood chips (or sawdust) are cooked in a digester under proper conditions of temperature and pressure by white liquor which is a solution of sodium hydroxide, sodium sulfide, sodium carbonate and sodium sulfite. The cooking process extracts the lignin from the wood leaving pulp. This pulp is further treated and converted to paper products. The cooking process also generates black liquor (lignin, impurities and remnants of the cooking solution). The kraft recovery cycle then converts this black liquor back to clear white liquor.

The recovery process of a kraft paper mill consists primarily of a liquor cycle where black liquor is evaporated, combusted and dissolved in weak wash to form green liquor (mostly a solution of sodium sulfide and sodium carbonate); then this green liquor is clarified, slaked, causticized and clarified again to yield white liquor suitable for cooking chips. The liquor process has as a sub-loop the lime recovery cycle. In the lime recovery process, lime mud (primarily calcium carbonate) settles out in the white liquor clarifier. The lime mud is removed from the clarifier as under flow. It is washed to remove excess soda, filtered to remove the wash water and then dried, heated and calcined in a rotary lime kiln to form lime (calcium oxide). This lime is then added to the green liquor in the slaker where it slakes. The green liquor clarification, slaking, causticizing, white liquor clarification and the lime recovery cycle in a kraft paper mill is called the recausticizing process.

The principle chemical reactions occurring in the recausticizing process are as follows:

(i) slaking: $CaO + H_2O \rightarrow Ca(OH)_2$,

(ii) causticizing: $Ca(OH)_2 + Na_2CO_3 \rightleftarrows 2NaOH + CaCO_3\downarrow$

(iii) calcining: $CaCO_3 \rightarrow CaO + CO_2\uparrow$

One obvious goal of the recausticizing process is to cause reaction (ii) to produce as much NaOH as possible and still retain economies in the recovery process.

The most important component in the lime recovery cycle is the rotary lime kiln in which the processes of evaporation, heat transfer, calcination and combustion occur and cause lime mud to become lime. This reburnt lime from the kiln should ideally be porous and completely calcined so that it will slake easily and its lime mud will settle rapidly. Such an ideal reburnt lime is produced when the dried lime mud is heated to 1550 - 1600°F and held there until all the CO_2 is evolved. If the temperature is too high, some of the lime becomes non-porous and difficulty in slaking occurs. If the temperature is too low and/or the lime has not burnt long enough, then some lime is not calcined and will not slake.

The burning of lime at 1550 - 1600°F (soft burning) is not feasible due to the length of time necessary for complete calcination to occur. Generally, the longer the rotary kiln the softer reburnt lime it produces but it is also more expensive to build than a shorter kiln. The standard rotary lime kiln length is presently thirty (30) times its

inner diameter, the diameter of the rotary kiln being dictated by the load the kiln is designed to carry.

Physically, a rotary lime kiln is cylindrical in shape and slightly sloped (1/2 inch per lineal foot is typical). The lime mud is fed into the upper end of the kiln and progresses slowly down the kiln (2'/min.is typical) as the lower end of the kiln the fuel (usually CH_4) is combusted and the hot gases pass over the burden material transferring heat to the burden material.

2. THE MATHEMATICAL MODEL

Although a dynamic model of a rotary lime kiln was not found in the literature, there are dynamic models of rotary cement kilns (1 and 3). As the processes of evaporation, heat transfer, calcination and combustion also take place in a rotary cement kiln, then the form of the mathematical model of a rotary lime kiln can be easily obtained from that of a rotary cement kiln.

Other papers have been published which deal with investigation of heat transfer in rotary kilns (4,5,7 and 8). Scatterfield et al. (6) investigated the calcining rate of calcium carbonate to determine when the reaction rate is controlled by the rate of mass transfer of CO_2, heat transfer and/or nucleation. Lyons et al. (2) present a steady state model of a rotary cement kiln which they used to study the effect of feed moisture on fuel economy. The dynamic model presented by Phillips (3) is rather simplified but his article

does include a good discussion of the differences in combustion of coal, oil and gas in a rotary kiln. The model in the paper by Min et al. (1) is the dynamic version of the model given in Reference (2).

The form of the mathematical equations describing the dynamic behaviour of a lime kiln consists of a system of first-order partial differential equations, where time, t, and longitudinal distance from feed end of kiln, l, are the independent variables. The equations are derived by performing mass and heat balances for the material in the kiln. The derivation and the resultant equations are given in Appendix A. The form of the equations is as follows:

$$\frac{\partial q(l,t)}{\partial t} = f(q,l,u, \frac{\partial q(l,t)}{\partial l}) \qquad (1)$$

where $t \in (t_o,t_f)$ some time interval, $l \in (0,L)$, q is an n-dimensional state vector, f( ) is an n-dimensional function and u is an m-dimensional vector of inputs.

To obtain the steady state model of the rotary lime kiln set $\partial q/\partial t$ equal to zero, then Equation (1) becomes

$$f(q,l,u, \frac{\partial q(l,t)}{\partial l}) = 0 \qquad (2)$$

In the rotary lime kiln model Equation (2) can be written in the form

$$\frac{dq(l,t)}{dl} = g(q,l,u) \qquad (3)$$

3. DESCRIPTION OF A KILN

The kiln for which data was collected during July and August of 1970 is located in the recovery section of the American Can Co. tissue plant at Halsey, Oregon.

The rotary kiln was made by Allis-Chalmers and installed in early '69. The kiln shell is 250' long and has an outside diameter of 9' -2 1/2" when cold and weighs about 560 tons. The kiln slopes toward the hot end at 1/2" per lineal foot. It is supported at four places where it turns on trunions at rates of 1.5, 1.125, 0.75, and 0.5625 rpm normally and at 0.1 rpm when the kiln is on auxiliary drive.

The interior of the kiln consists of three sections: (i) chain section, (ii) preheating section and (iii) combustion zone. The drying zone extends from 0' to 34' and is lined with castable refractory to a 4 1/2" thickness. The chains are hung for 26' from 6' to 32'. The preheating section (34' to 170') is lined with 6" thickness 40% alumina refractory brick and mortar. The combustion zone (170' to 250') is lined with a 6" thickness of 70% alumina brick and mortar.

4. MEASUREMENTS

The measurements available from the kiln are shown in Table I and the points at which these measurements were made are shown in Figure 2.

The measurements of the components of the burden mass at points p_1, p_2, p_3, and p_5 are obtained by grabbing a sample and having it analyzed. The gas temperatures at p_1, p_3, and p_4 are measured by thermocouples and the burden temperature at p_5 is measured by means of an optical pyrometer. At p_1, p_2 and p_3, burden temperature is determined by putting a thermometer into the sample immediately after sampling. The gas components are measured by grabbing a sample and analyzing.

Flow rate can be measured by using a pitot tube.

The measurement of the burden temperature at p_5 is sometimes questionable due to dusting. The calibration of the temperature sensors was not checked. The gas temperature at p_5 is an estimate as is the amount of water vapor at p_1. Also flame length is difficult to judge anyway but in the kiln of Section 3 the viewing port is on the right and low so that the view into the kiln is obscured by the burden in the kiln. The actual rate of lime feed into the kiln cannot be measured directly for the kiln of Section 3 and only the flow rate to the de-watering centrifuge is available.

The typical operating conditions of the kiln of Section 3 are given in Table II.

5. METHOD OF PARAMETER ESTIMATION

The steady state kiln equations and the measurements equations are given as follows:

Plant: $$\frac{dq}{dl} = F(q,u,l,\alpha),\ l \in (0,L) \qquad (4)$$

Measurements: $$z(p_i) = H(p_i)q(p_i)\ ,\ i=1,\ \ldots\ ,\ 5 \qquad (5)$$

where q is the 8th order state vector, u is a p-dimensional vector of known inputs, $F(\ )$ is an 8-dimensional vector-valued function, α is a vector of unknown parameter values and H is an 8 x 8 diagonal matrix.

A performance index is chosen

$$J(\alpha) = \sum_{i=1}^{5} \| z(p_i) - H(p_i)\hat{q}(p_i,\alpha) \|_W^2 \qquad (6)$$

where W is an 8x8 weighting matrix and $\hat{q}$ is a solution of the

plant. Equation (4) for parameter values equal to α.
Problem: find a value of $\alpha = \alpha^{o}$ such that $J(\alpha)$ is minimal.
Solution: choose an initial value of $\alpha = (\alpha_{init.})$ and iterate on α using a gradient technique.

Many gradient techniques are available, and these are now being simulated. The major problem with the gradient method in this instance is the non-linearity of the lime kiln equations. It is noted that gradient technique could determine only a local minimum.

TABLE I

MEASUREMENTS ON KILN

Measurement Point	Quantities Measured
p_1	$CaCO_3$, CaO, $H_2O(l)$, $H_2O(g)$, O_2, CO_2, T_s, T_g
p_2	$CaCO_3$, CaO, $H_2O(l)$, T_s
p_3	$CaCO_3$, CaO, $H_2O(l)$, T_s, T_g
p_4	T_g
p_5	$CaCO_3$, CaO, $H_2O(g)$, CH_4, O_2, CO_2, T_s, T_g, $H_2O(l)$

TABLE II

KILN DATA

Burden mass temperature at p_1 (°F)	115
" " " " p_2 "	160
" " " " p_3 "	180
" " " " p_5 "	2000
Gas temperature at p_1 (°F)	510
" " " p_3 "	1100
" " " p_4 "	1400
Specific heat of burden mass (Btu/lb. -°F)	0.26
" " " gas (Btu/lb. -°F)	0.28
Calcinable mass flow rate (lb./hr.)	5.0×10^3
Nitrogen " " " "	2.7×10^4
Fuel " " " "	1.3×10^3
Water at p_1 (%) (wet basis)	35
" " p_2 " "	16
" " p_3 " "	3
Oxygen at p_1 (%) (dry basis)	6.8 - 7.2
Carbon dioxide at p_1 (%) (dry basis)	14
Active * lime at p_5 (%)	89.6
Gas temperature at p_5 (°R) (est.)	750
Residence time (hr.)	1 1/2 - 2

* A.S.T.M. "rapid" sugar test was used to determine % active lime

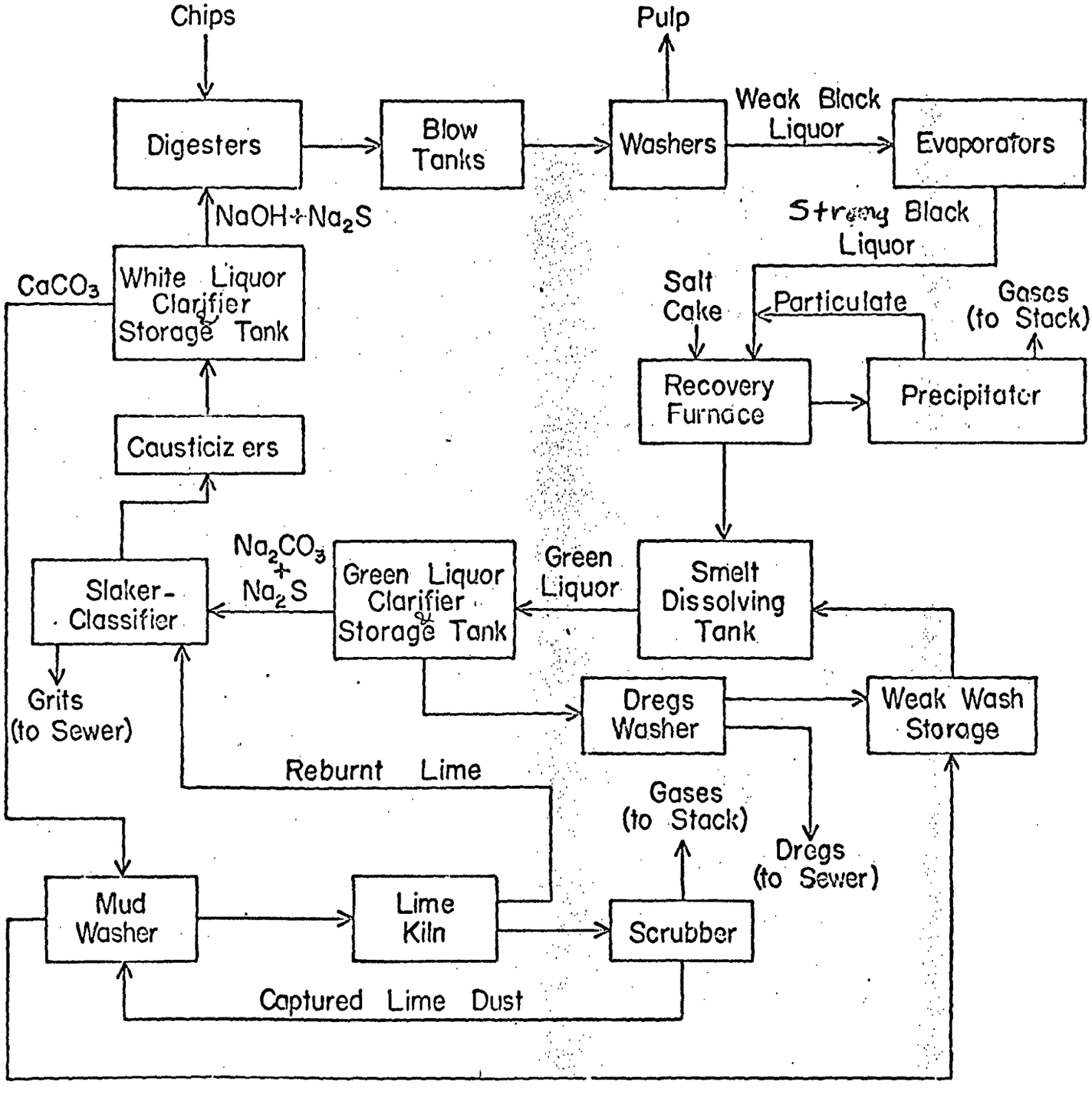

FIGURE 1. BLOCK DIAGRAM OF CHEMICAL RECOVERY PROCESS OF A KRAFT PAPER MILL.

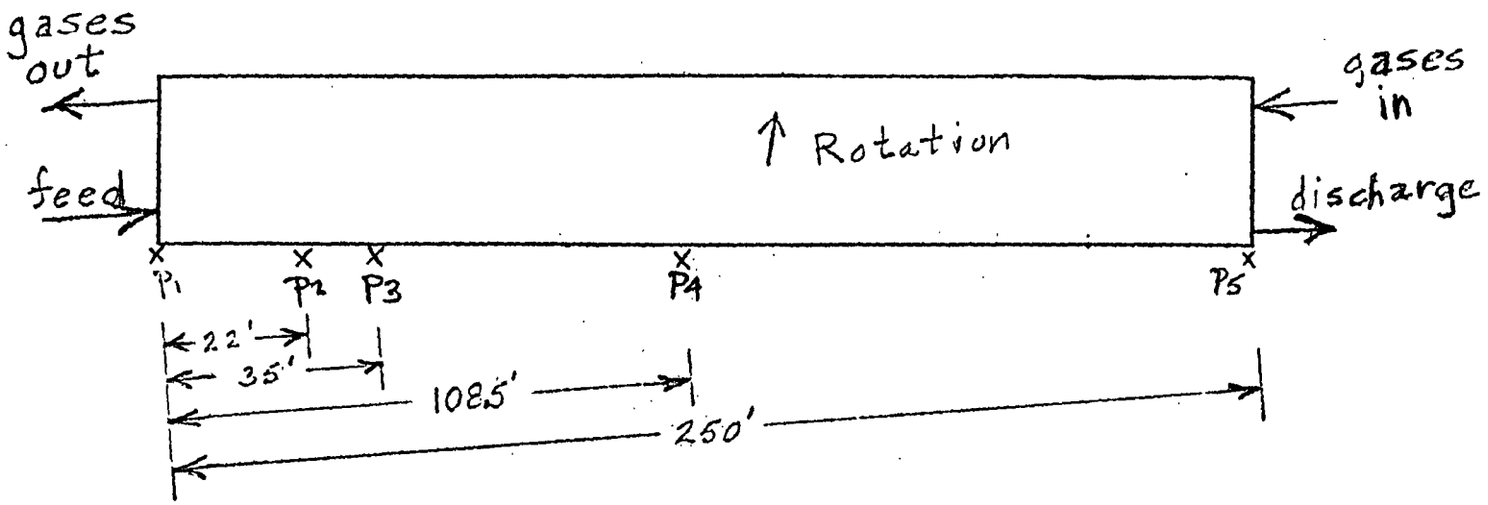

Fig. 2. Measurement Points along Kiln

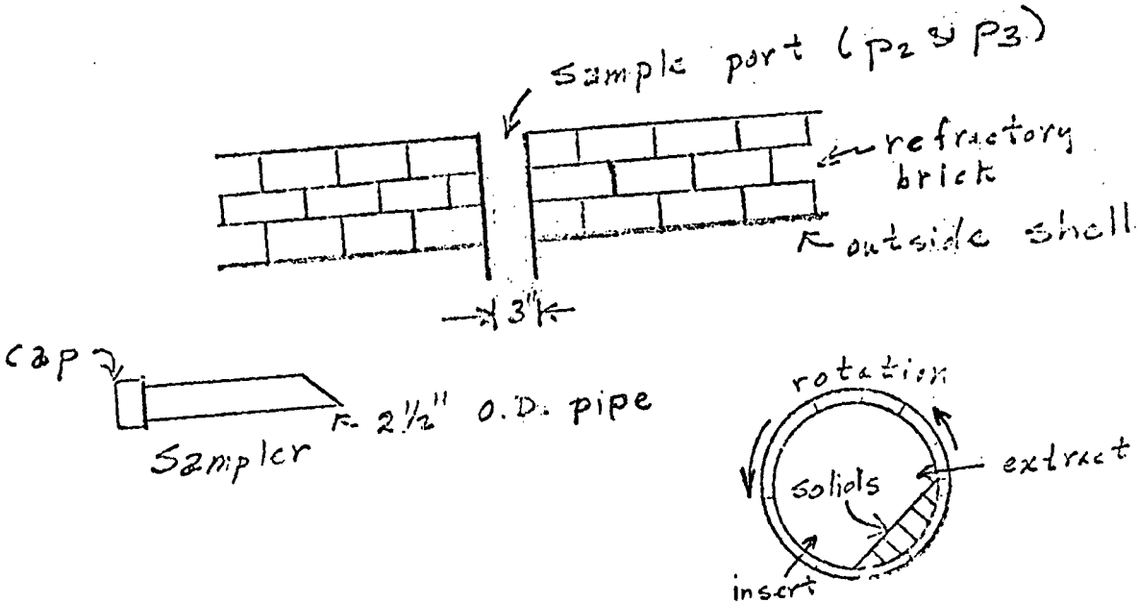

Fig. 3. Sampling Technique @ P2 & P3.

APPENDIX A

I. ASSUMPTIONS

These assumptions are used in the derivation of equations:

(i) The specific heats, latent heat and the heats of reaction are constant.

(ii) The fuel is methane (CH_4) and the combustion is complete.

(iii) Possible impurities and inert compounds are neglected in the material and energy balances.

(iv) Possible solid particles in the gas flow are neglected.

(v) The residence time of the burden mass in the kiln is inversely proportional to the angular speed and to the length of the kiln.

(vi) The burden mass held in the kiln at any time is directly proportional to the flow rate of the burden mass.

(vii) The temperature distribution of the kiln wall is independent of radial position.

II. DERIVATION OF THE EQUATIONS FOR A LIME KILN

The total rate of change of material Q at a point(ℓ,t) or the reaction rate is the sum of the rate of change of Q at a fixed value of ℓ and the rate of change of Q due to flow of the material Q. Hence,

$$\frac{\partial Q(\ell,t)}{\partial t} = \frac{d}{d\ell}\left[iQ(\ell,t)\right] - \frac{\partial Q(\ell,t)}{\partial \ell} \quad \frac{d\ell}{dt} \qquad (A1)$$

The term $d\ell/dt$ is specified for the burden flow by

$$d\ell/dt = W_b/B, \qquad (A2)$$

and for the gaseous flow by

$$\frac{d\ell}{dt} = \frac{W_{gN}}{N} \qquad (A3)$$

By the conservation of the material, the equations for the material balances in the kiln can be written using equations (A1,), (A2) and assumption (vi).
It results in equations (A4) through (A8):

CaO in the form of $CaCo_3$:

$$\frac{\partial X_{CaCO}(\ell,t)}{\partial t} = -K_C X_{CaCO_3}(\ell,t) - \frac{W_b}{B} \frac{\partial X_{CaCO_3}(\ell,t)}{\partial \ell} \qquad (A4)$$

CaO:

$$\frac{\partial X_{CaO}(\ell,t)}{\partial t} = -K_C X_{CaO}(\ell,t) - \frac{W_b}{B} \frac{\partial X_{CaO}(\ell,t)}{\partial \ell} \qquad (A5)$$

Water:

$$\frac{\partial X_W(\ell,t)}{\partial t} = -R_W - \frac{W_b}{B} \frac{\partial X_W(\ell,t)}{\partial \ell} \qquad (A6)$$

O_2:

$$\frac{\partial Y_{O_2}(\ell t)}{\partial t} = -4.0\, R_F + \frac{W_{gN}}{N} \frac{\partial Y_{O_2}}{\partial \ell} \qquad (A7)$$

Water Vapor:

$$\frac{\partial Y_W(\ell,t)}{\partial t} = 2.25\, R_F + \frac{B}{N} R_W + \frac{W_{gn}}{N} \frac{\partial Y_W(\ell,t)}{\partial \ell} \qquad (A8)$$

CO_2 :

$$\frac{\partial Y_{CO_2}(\ell,t)}{\partial t} = 2.75\, R_F + 0.786 \frac{B}{N} K_C X_{CaCO_3}(\ell,t) + \frac{W_{gN}}{N} \frac{\partial Y_{CO_2}(\ell,t)}{\partial \ell} \qquad (A9)$$

In equations (A4), (A5) and A6), the burden material is normalized with respect to B, the calcinable mass per

unit length. In equations (A7), (A8) and (A9), the gas material is normalized with respect to N, the nitrogen mass per unit length.

For example, consider the rate of change of water vapor Y_W at a point (ℓ,t). Then

$$\frac{\partial Y_W(\ell,t)}{\partial t} = \text{reaction rates} + \frac{W_{gN}}{N}\frac{\partial Y_W(\ell,t)}{\partial \ell}$$

$$2.25\,(\text{rate of combustion}) + \frac{B}{N}\,(\text{rate of evaporation of water}$$

$$+ \frac{E_{gn}}{N}\frac{\partial Y_W(\ell,t)}{\partial \ell} \qquad \text{(A10)}$$

where the factor 2.25 is the ratio of mass of water vapor formed per unit mass of CH_4 combusted. The factor B/N appears because of multiplying by B makes the term an absolute quantity and dividing by N normalizes it with respect to pounds of nitrogen/unit foot.

On the basis of the energy balance, the following equations are written for the burden temperature:

$$Cp_S M_S \frac{\partial T_S(\ell,t)}{\partial t} = h_2A_2(T_g-T_S) + h_3A_3(T_W-T_S) - BH_WR_W - BH_CK_CX_{CaCO_3} - Cp_SW_S\frac{\partial T_S(\ell,t)}{\partial \ell} \qquad \text{(A11)}$$

- gas temperature:

$$Cp_gM_g\frac{\partial T_g(\ell,t)}{\partial t} = h_1A_1(T_W-T_g) + h_2A_2(T_S-T_g) + NH_FR_F - Cp_S(T_S-T_g)B(-0.786K_CX_{CaCO_3}-R_W) + Cp_gW_g\frac{\partial T_g(\ell,t)}{\partial \ell} \qquad \text{(A 12)}$$

- temperatures of the walls inside and outside

$$\left.\begin{aligned} h_1A_1(T_g-T_W) + h_3A_3(T_S-T_W) + h_4A_4(T_W'-T_W) &= 0 \\ h_4A_4(T_W-T_W1) + h_5A_5(T_a-T_W1) &= 0 \end{aligned}\right\} \qquad \text{(A 13)}$$

As an example, equation (A11) is derived. The equations stating the conservation of the energy of the gas can be written since the rate of change(per hour) of the heat energy in the material $\frac{d}{dt}[Cp_m M_m(\ell,t]$ is equal to the sum of (i) the rates of heat transfer into the material from the wall and gases or burden mass, (ii) the heat rate of the reactions (combustion, calcining, and evaporation), and (iii) the heat rate from the addition of material at temperature T_m' due to changes of state.

For example, the heat energy balance for the stream is as follows:

$$\frac{d}{dt}\left[Cp_g M_g(\ell,t)T_g(\ell,t)\right]$$

$$= Cp_g M_g\left(\frac{\partial T_g(\ell,t)}{\partial t} + \frac{\partial T_g(\ell,t)}{\partial \ell}\ \frac{d\ell}{dt}\right) + Cp_g T_g\left(\frac{\partial M_g(\ell,t)}{\partial t}\right)$$

$$+ \frac{\partial M_g(\ell,t)}{\partial \ell}\ \frac{d\ell}{dt}\ \Big) = Cp_g M_g \frac{\partial T_g(\ell,t)}{\partial t} - Cp_g W_{gN}\frac{\partial T_g(\ell,t)}{\partial \ell}$$

$$- Cp_g T_g\left(\frac{\partial M_S(\ell,t)}{\partial t} + \frac{\partial W_b(\ell,t)}{\partial \ell}\right) = h_1 A_1(T_W - T_g)$$

$$+ h_2 A_2(T_S - T_g) + NH_F R_F - Cp_g T_S\left(\frac{\partial W_S(\ell.t)}{\partial \ell} + \frac{\partial M_S(\ell,t)}{\partial t}\right) \qquad (A\ 14)$$

The last two terms in equation (A14) account for gases, water vapor and carbon dioxide, which is added to the gas flow due to evaporation and calcining, respectively, Equation (A11) is then obtained by rearranging the terms in equation (A14).

In addition to equations (A4) through (A13) the following constraints must be satisfied:

- The mass flow rate of the burden is equal to:

$$W_S = W_b(1+0.786\ X_{CaCO_3} + X_W) \qquad (A15)$$

- The total burden mass per unit foot is equal to:

$$M_S = B(1 + 0.786\ X_{CaCO_3} + X_W) \qquad (A16)$$

- The mass flow rate of the gas is:

$$W_g = W_{gN}(1 + Y_{O_2} + Y_W + Y_{CO_2} + Y_F) \qquad (A17)$$

- The total mass of the gas is expressed by:

$$M_g = N(1.0 + Y_{O_2} + Y_W + Y_{CO_2} + Y_F) \qquad (A18)$$

The heat transfer rates (B.t.u./(hr.)(ft.^{2}/ft.) (^{0}R) are written when the transfer occurs from

-gas to wall: (A19)

$$h_1 = f_1 + 1.73 \times 10^{-9}\ \epsilon_g\epsilon_S\ (T_g^{\ 3} + T_g^{\ 2}T_S + T_gT_S^{\ 2} + T_S^{\ 3})$$

- gas to the burden:

$$h_2 = f_2 + 1.73 \times 10^{-9}\ \epsilon_g\epsilon_S(T_W^{\ 3}+T_g^{\ 2}T_S+T_gT_S^{\ 2}+T_S^{\ 3}) \qquad (A20)$$

- The inner wall to the burden:

$$h_3 = f_3\, 1.73 \times 10^{-9}\left(\frac{A_2}{A_3}\right)\epsilon_W\epsilon_S(T_W^{\ 3}+T_W^{\ 2}T_S+T_WT_S^{\ 2}+T_S^{\ 3}) \qquad (A21)$$

- The inner wall to the outer wall of the kiln:

$$h_4 = f_4 \qquad (A22)$$

- The outer wall to the surrounding medium:

$$h_5 = f_5 + 1.73 \times 10^{-9}\epsilon_W(T_W'^{\ 3}+T_W'^{\ 2}T_a+T_W'T_A^{\ 2}+T_a^{\ 3}) \qquad (A23)$$

Rates and rate constants are as follows:

- The reaction rates:

$$K_C = A_C \exp\left[-\Delta E_C/RT_S\right] \tag{A24}$$

$$K_W = A_W \exp\left[-\Delta E_W/RT_S\right] \tag{A25}$$

- The drying rate:

$$R_W = \begin{cases} K_W & \text{for } X_W \geq 0.1 \\ K_W X_Y & \text{for } X_W \leq 0.1 \end{cases} \tag{A26}$$

- The combustion rate:[(lb/lb of H_2)/hr]: (A27)

$$R_F = \frac{Y_F}{L-200} (W_{gN}/N), \quad \ell > 200$$

In equation (A23), a uniform burning in combustion zone has been assumed.

The speed of the burden is defined by

$$\frac{\partial L}{\partial \theta} = \frac{Wb}{B} \tag{A28}$$

Equation (A4) through (A13) and (A15) through (A28) establish the mathematical form which represents the behavior of the kiln.

APPENDIX B

NOMENCLATURE

A = heat transfer area (sq. ft./ft.)

B = calcinable mass/unit length (lb./ft.)

C_{p_g} = specific heat of gas (B.t.U./lb.-°R)

C_{p_S} = specific heat of burden (B.t.U./lb.-°R)

ϵ = emissivity

ΔE = activation energy (B.t.u./lb. -mole)

f = conductive and convective heat transfer coefficient (B.t.u./hr. - sq. ft. - °R)

h = heat transfer coefficient (B.t.u./hr.-ft. - °R)

H = heat of reaction (B.t.u./lb.)

K = reaction rate coefficient ($hr.^{-1}$)

L = length of kiln (ft.)

M = molecular weight

M_g = mass of gas/unit length (lb./ft.)

M_S = burden mass/unit length (lb/ft.)

N = Nitrogen mass/unit length (lb./ft.)

RF = rate of combustion/unit nitrogen mass (lb.-hr.)

R_W = rate of evaporation/unit calcinable mass (lb./16.-hr.)

θ = residence time (hr.)

t = real time (hr.)

T_a = ambient temperature (°R)

T_g = gas temperature (°R)

T_S = burden temperature (°R)

T_W = inner wall temperature (°R)

T'_W = outer wall temperature (°R)

W_b = calcinable mass flow rate (lb./hr.)

APPENDIX B (cont.)

W_g = gas mass flow rate (lb./hr.)

W_{gN} = nitrogen mass flow rate (lb./hr.)

W_S = burden mass flow rate (lb./hr.)

X_{CaCO_3} = CaO in form of $CaCO_3$/unit calcinable mass (lb./lb.)

X_{CaO} = free CaO/unit calcinable mass (lb./lb.)

X_W = water/unit calcinable mass (lb./lb.)

Y_{O_2} = oxygen/unit nitrogen mass (lb./lb.)

Y_W = water vapor/unit nitrogen mass (lb./lb.)

Y_{CO_2} = carbon dioxide/unit nitrogen mass (lb/lb.)

Y_F = fuel/unit nitrogen mass (lb./lb.)

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OVERALL STEADY-STATE MODEL OF THE KRAFT RECOVERY CYCLE

BY

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The steady-state program can be separated into two main areas: (1) The development of a mathematical model describing the steady-state conditions of the Kraft Recovery Cycle and (2) the minimization of a performance index involving the control set-point and output variables with the mathematical model as a constraint. The performance index will be expressed in terms of cost.

As stated in the previous report, The Kraft Process comprises five areas or component systems:

1) Digester and washers
2) Multi-effect evaporators
3) Recovery furnace, cascade evaporator and associated equipment
4) Clarification and causticizing equipment
5) Lime kiln and associated calcination equipment

This past summer was spent at Consolidated Papers, Inc. in order to become acquainted with the Kraft Process. One of the objectives was to identify the independent variables which could be used to control the operation and also determine the additional dependent variables required to develop the steady-state model. The following list is the variable definitions for the evaporator and furnace systems; classification of variables for the remainder of the Kraft Cycle will be completed shortly.

I. EVAPORATOR SYSTEM

A. Controls

U_1 = Steam flow to 1st effect

U_2 = Pressure in last effect

B. Outputs

Y_1 = Black liquor flow from 1st effect

Y_2 = Pollutants in vapor from last effect

Y_3 = Condensate flow from system

Y_4 = Soap from skim tank

C. Disturbances

d_1 = Steam pressure to 1st effect

d_2 = Water temperature to surface condenser

d_3 = Ambient temperature

d_4 = Tube fouling

d_5 = Weak black liquor flow to last effect

II. RECOVERY SYSTEM

A. Controls

U_1 = Black liquor flow to cascade evaporator

U_2 = Liquor bath level in cascade

U_3 = Salt cake addition

U_4 = Black liquor temperature to furnace

U_5 = Primary air flow to furnace

U_6 = Secondary air flow to furnace

U_7 = Primary/secondary air temperature

B. Outputs

Y_1 = Smelt flow

Y_2 = Steam production rate

Y_3 = Pollutants

C. Disturbances

d_1 = Black liquor solids flow to furnace

d_2 = Change in heat transfer rate to furnace boiler system

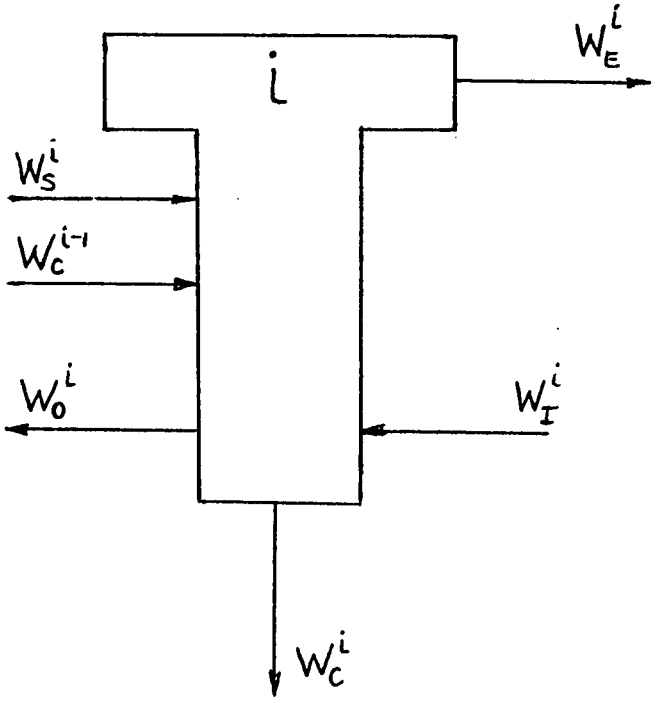
d_3 = Precipitator efficiency

d_4 = Change in ambient conditions

The present status of the overall steady-state model is the completion of the multi-effect evaporator system. The model is based on two mass and two energy balances for each effect and include several important assumptions which will be discussed shortly.

A list of variable definitions is included at the end of this report.

Consider the following lumped-model for an $i^{\underline{th}}$ effect:



A mass balance on the steam and condensate can be written as

$$W_S^i + W_C^{i-1} = W_C^i \qquad (1)$$

Likewise, an energy balance can also be written as

$$W_S^i H_S^i + W_C^{i-1} = h^i A^i (T_S^i - T_O^i) + W_C^i H_C^i \qquad (2)$$

In a similar manner a mass balance on the black liquor can be described as

$$W_I^i = W_O^i + W_E^i \qquad (3)$$

And the energy balance is

$$W_I^i C_{P_{BLI}}^i T_O^i + W_S^i H_S^i = W_O^i C_P^i T_O^i + W_E^i + H_E^i \qquad (4)$$

An additional relation is

$$T_O^i = T_{SAT}^i + BPR\ (X_I^i) \qquad (5)$$

These five equations serve a two-fold purpose:

(1) To predict the outputs, given the inputs,

(2) To give a time history of the effect of fouling in terms of a heat transfer coefficient for each effect.

The primary assumption involved is that all the energy release by the totally condensed steam is transferred to the black liquor.

The procedure for using the model would be to calculate a set of h's from a standard set of inlet steam and black liquor conditions on a daily basis and then to use those h values throughout the day to predict outputs for input changes.

DEFINITION OF VARIABLES

A. FLOW RATES (lbm/hr)

W_C^i = Condensate from $i^{\underline{th}}$ effect

W_E^i = Evaporated water from $i^{\underline{th}}$ effect

W_I^i = Black liquor input to $i^{\underline{th}}$ effect

W_O^i = Black liquor output from $i^{\underline{th}}$ effect

W_S^i = Steam input to $i^{\underline{th}}$ effect

(NOTE: $W_S^i = W_E^{i-1}$ for $i \neq 1$)

B. PHYSICAL PROPERTIES

$C_{P_{BL}}$ = Specific heat of black liquor (Btu/lbm $ft^2$of)

X^i = % Solids (by weight) of black liquor to $i^{\underline{th}}$ effect (decimal fraction)

H_C^i = Enthalopy of condensate from $i^{\underline{th}}$ effect (btu/lbm)

H_S^i = Enthalopy of steam TD $i^{\underline{th}}$ effect (btu/lbm)

C. TEMPERATURES (°F)

T_I^i = Inlet temperature of black liquor TD $i^{\underline{th}}$ effect

T_O^i = Outlet temperature of black liquor from $i^{\underline{th}}$ effect

BPR = Boiling point rise

D. PHYSICAL PARAMETERS

A^i = Heat transfer area of $i^{\underline{th}}$ effect (ft^2)

h^i = Heat transfer coefficient of $i^{\underline{th}}$ effect (Btu/ft^2°F hr)

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